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DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION (SCR)
TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO$_X$)
EMISSIONS FROM HIGH-SULFUR COAL-FIRED BOILERS

Final Report
(Appendices O-T)
Volume 3 of 3

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TESTING AND ANALYTICAL SERVICES FOR THE INNOVATIVE CLEAN COAL TECHNOLOGY DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION (SCR) TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO_x) EMISSIONS FROM HIGH SULFUR COAL

Final Report for Task 4

Air Heater Tests
Reactors A, B, C
Second Parametric Series

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TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

ACKNOWLEDGMENTS

Section 1

INTRODUCTION

This report is the eighth in a series of reports describing the results of testing and analytical services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The test results for the second sequence of air heater tests are summarized in this report. The air heater evaluations were conducted at the beginning (June 1994) and end (October 1994) of the second sequence of parametric testing under Task 4: Long-Term Parametric Tests. The tests in June 1994 (SO_2, SO_3, and particulate mass concentrations) were specifically conducted to establish a set of baseline performance values. (The air heaters had been washed and a partial replacement of baskets had occurred on the two rotary air heaters (A and B) during the May/June 1994 outage.)

The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The test facility includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 °F down to 300 °F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C's air heater incorporates a heat pipe design.

The original design of the SCR test facility also included bypass heat exchangers on each of the three large reactors. These units were intended for use during parametric testing of the reactors (Task 4: Long Term Parametric Tests) so that flue gas containing higher concentrations of slip ammonia could be diverted around the air heaters. However, the bypass heat exchangers did not function as planned and the practice of bypassing the air heaters on Reactors A, B and C during parametric testing was abandoned. Thus, the air heater test data presented in this report include the effects of day-long periods of exposure to ammonia concentrations normally ranging from a few parts per million by volume ppm(v) to 20 ppm(v), although brief excursions approached 100 ppm(v).

The air heater testing included the determinations of particulate mass concentration (outlet of Reactor A and B air heaters) and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia (inlet and outlet of Reactor A, B and C air heaters). All tests were conducted with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. In general, simultaneous tests were performed at the inlet and outlet of each of the three air heaters.

This report is divided into several sections. Section 2 describes the test methods used for the air heater testing. Section 3 reviews all of the test results. The test data are briefly summarized in Section 4. Tables containing original data summaries that were produced for the air heater testing are contained in Appendix A.

Section 2

TEST METHODS

A variety of test methods were used to characterize air heater performance. The following subsections briefly describe the test methods for ammonia (NH_3), sulfur dioxide (SO_2), sulfur trioxide (SO_3), hydrogen chloride (HCl), and particulate mass concentration. Figure 2-1 shows a sketch of a large reactor and its air heater (not to scale). The test port locations for air heater inlet and outlet measurements are indicated on the diagram. Test ports (three horizontal ports) downstream of the third catalyst layer (the normal reactor outlet test location) were used to measure concentrations of air heater inlet gas constituents. Three test ports (horizontal) were installed in a transition piece in the outlet ducting of the Reactor A and B air heaters that were suitable for particulate and gas phase flue gas constituent testing. Test ports at this same location in the Reactor C air heater outlet were not installed. However, a single test port (horizontal) in a section of ducting at the inlet transition to the cyclone on Reactor C (downstream of the reactor bypass duct) was suitable for testing gas phase flue gas constituents (Reactor C air heater outlet). Turbulent flow at this location precluded mass concentration tests, however.

Air heater tests were performed with the reactors operating at Test Condition 22, the normal baseline or long-term operating condition. These operating conditions included a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 0.8, and a flue gas flow rate of 5000 wscfm (100% of design flow rate). During measurements of air heater inlet and outlet ammonia concentrations only, additional tests were performed at Test Condition 24. The operating parameters for these tests were a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 1.0, and a flue gas flow rate of 5000 wscfm.

AMMONIA

Ammonia concentrations were measured simultaneously at the air heater inlet (downstream of the third catalyst layer) and at the air heater outlet on each of the large reactors. At each site and at each test condition, three independent tests were performed to determine an average ammonia concentration. Sampling ports located downstream of the third catalyst layer were used to measure air heater inlet ammonia concentration. Flue gas was sampled at three equally-spaced points in each of the three test ports (a total of nine traverse points) to determine an average inlet ammonia concentration for each test. At the air heater outlet sampling location on Reactors A and B the probe also sampled at three equally-spaced points in each of the three test ports (a total of nine traverse points). At the single Reactor C outlet test port, the ammonia probe traversed three, equally-spaced, points (front to rear) during each test.

LOCATION OF AIR HEATER INLET TEST PORTS (THREE HORIZONTAL) ON REACTORS A, B, C

LOCATION OF 4th CATALYST LAYER

(EMPTY)

GAS FLOW

NOT TO SCALE

AIR HEATER SECTION

LOCATION OF AIR HEATER OUTLET TEST PORTS (THREE HORIZONTAL) ON REACTORS A & B

REACTOR BYPASS DUCT

LOCATION OF AIR HEATER OUTLET TEST PORT (ONE HORIZONTAL) ON REACTOR C (GAS PHASE ONLY)
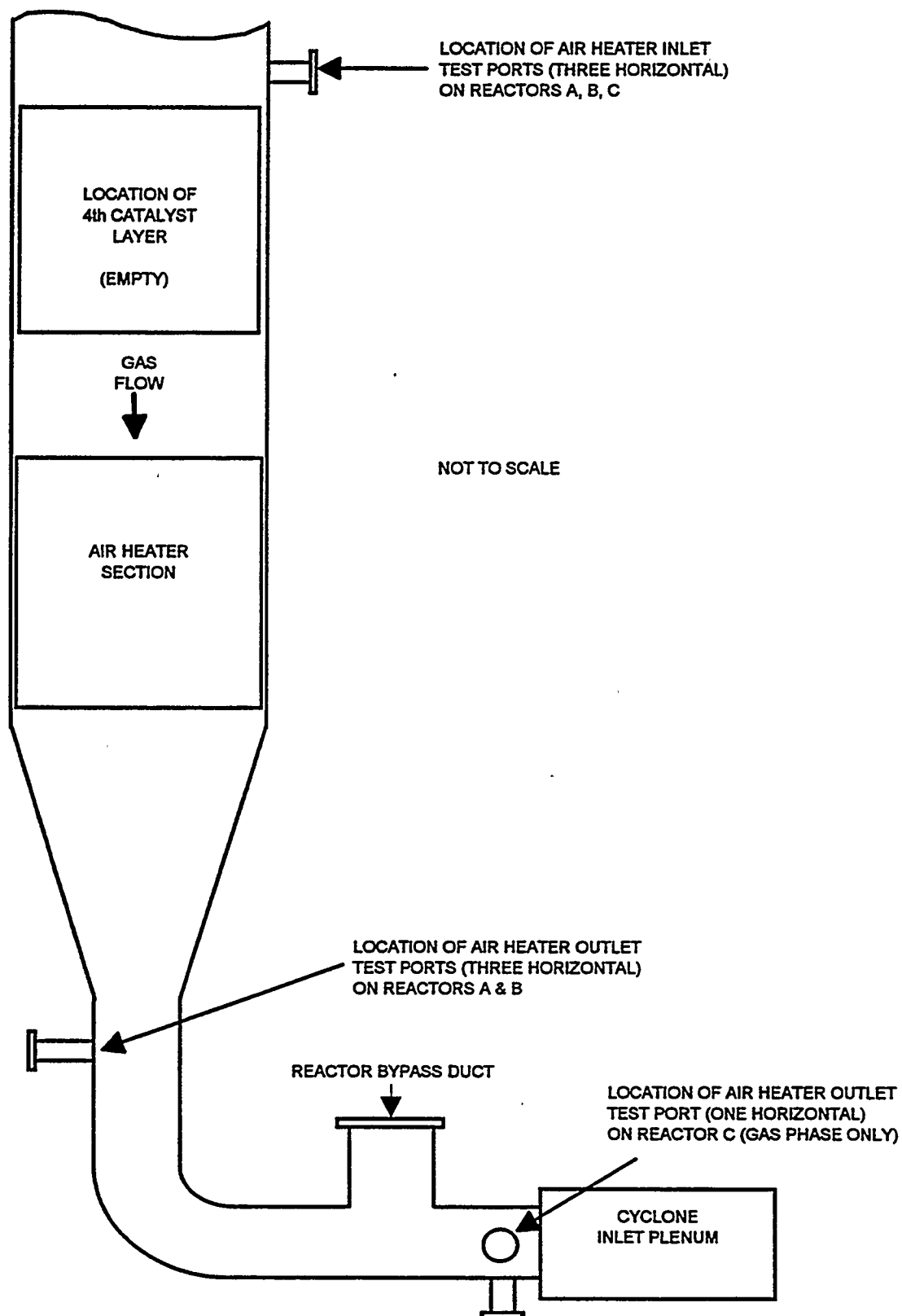
CYCLONE INLET PLENUM

Figure 2-1. Schematic drawing of SCR reactor and air heater cross section (side view).

Figure 2-2. Schematic diagram of the ammonia sampling train.

A schematic drawing of the ammonia sampling train is shown in Figure 2-2. The flue gas sample is drawn through a glass-lined probe and a heated filter to remove particulate. The filter is maintained at the nominal flue gas temperature. The gas sample passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 0.1 normal solution of sulfuric acid. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. A second glass wool plug is placed in the line between the last two impingers to collect any ammonia aerosols that may escape the second impinger. The remainder of the train consists of a silica gel column to remove the last traces of water from the flue gas sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The samples were segregated during sample collection into a solid-phase sample and a gas-phase sample. The solid-phase sample consisted of the heated filter, all of the collected particulate, and the probe-wash liquid. The gas-phase sample consisted of the impinger liquids and the wash and rinse liquids. The two samples were analyzed separately to characterize the ammonia partitioning between the gas and solid phases.

At the outset of the Task 4 parametric testing, modified Greenberg-Smith type impingers were used for ammonia sampling. However, 100 ml gas sampling impingers, each containing about 50 ml of solution, were used for the air heater ammonia testing to reduce the detection limit to less than 1 ppm(v) for a reasonable sample volume, about 3 ft^3 of gas. The lower detection limit was needed for much of the air heater ammonia testing because ammonia concentrations are very low at the reactor exit when operating at the reactor design condition with an NH_3/NO_x ratio of 0.8.

The impinger solutions are made alkaline in the laboratory (converting the NH_4^+ ion to free NH_3 in solution). The concentration of ammonia is then determined with an ammonia ion specific electrode, Orion Model 920A. As mentioned above, three individual measurements of ammonia concentration are made for each test condition. Two independent determinations of the ammonia concentration are then conducted on each sample.

SULFUR DIOXIDE AND SULFUR TRIOXIDE

During the air heater evaluation, SO_2 and SO_3 concentrations were measured simultaneously at the reactor outlet downstream of the third catalyst layer (air heater inlet) and the air heater outlet. Sulfur dioxide (SO_2) and sulfur trioxide (SO_3) were collected in a controlled condensation

sampling train. All tests were conducted using single-point sampling in either the right-hand port (air heater inlet test location) or middle port (air heater outlet test location).

Figure 2-3. Schematic diagram of the controlled condensation sampling train for SO_2 and SO_3.

A schematic drawing of the SO_2/SO_3 sampling train is shown in Figure 2-3. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The next element in the train is the SO_3 condenser. The condenser is a length of quartz tubing packed with quartz wool and maintained between 120 °F and 130 °F in a heated water bath. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 3% solution of hydrogen peroxide to oxidize SO_2. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The SO_3 is collected in the condensing element and the SO_2 is collected in the bubblers by oxidation with the hydrogen peroxide, converting it to H_2SO_4 in solution. In the condenser the SO_3 present begins a hydration reaction with the water vapor present making H_2SO_4. The excess water vapor also condenses to produce a condensate of concentrated aqueous H_2SO_4. Thus, two solutions of H_2SO_4 are collected; one a very concentrated solution of limited amount containing the original SO_3 and the other a relatively weak solution in far greater amount containing the original SO_2. The concentrations of the sulfate ion are determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph.
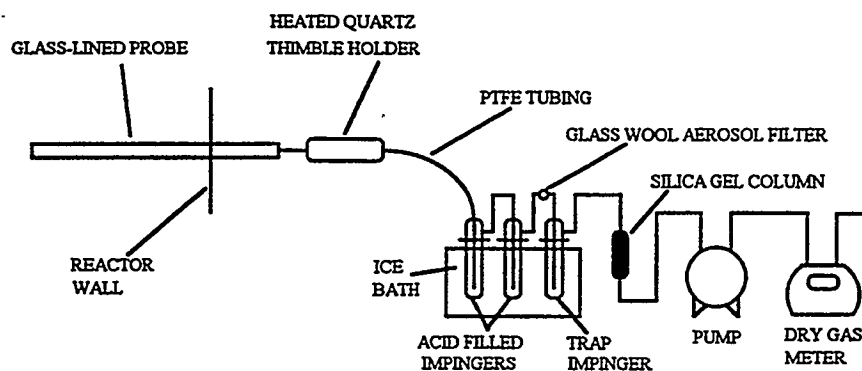
HYDROGEN CHLORIDE

Figure 2-4. Schematic diagram of the chloride sampling train.

A schematic sketch of the chloride sampling train is shown in Figure 2-4. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain type 1 de-ionized water for removal of HCl vapor. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The chloride ion concentration is determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph. Hydrogen chloride concentrations were determined by single-point sampling (at the mid-point of the duct) at the air heater inlet (downstream of the third catalyst layer, right-hand port) and at the air heater outlet (middle port). Inlet and outlet tests were conducted simultaneously.

MASS CONCENTRATION

Mass concentration was measured using a sampling train similar to that defined in the EPA Reference Test Method 17. A schematic drawing of the train used for mass concentration sampling is shown below in Figure 2-5.

Figure 2-5. Schematic drawing of the mass concentration sampling train.

This train consists of a filter holder and nozzle that is designed to operate in the flue gas stream, followed by a heated sampling probe, a condenser, a drying column, a gas meter, a pump, and finally a flow control mechanism, usually a calibrated orifice. An S-type pitot and thermocouple located near the nozzle provide a means for sampling isokinetically during each test. The glass fiber thimbles are desiccated before and after sampling and then weighed on an electronic microbalance.

Particulate mass concentrations were measured simultaneously upstream of catalyst layer 4 (air heater inlet) and at each air heater outlet (except Reactor C). At the air heater inlet, the three sample ports immediately above the access door at the fourth catalyst layer were used to obtain the sample. During each test particulate was sampled at three, equally-spaced, positions within the three test ports (a total of nine traverse points). At the air heater outlets on reactors A and B, three ports are available for mass sampling. During each test particulate was sampled at three, equally-

2-6

spaced, positions within the three test ports (a total of nine traverse points). Three runs were made at each test site to give a meaningful statistical average.

.SAMPLING QUALITY ASSURANCE

The measures adopted to ensure that meaningful results were obtained during the various testing procedures can be divided into three categories; equipment maintenance and calibrations, operating techniques, and analytical techniques. New equipment was obtained for use in the SCR testing program and preventive maintenance and calibrations are performed at regular intervals. Due to the dynamic nature of the process, each measurement requires three replicate samples. The replicates are averaged to yield a representative value. Variability gives a means to discern any anomalies not revealed by other quality control checks.

The sample trains are leak-checked by drawing a vacuum of 15 in. Hg before and after each sampling run. No in-leakage is tolerated. If any loss of pressure is observed within one minute, the source of the leakage is found and eliminated prior to testing. During each run the oxygen level of the gas exiting the dry gas meter is measured to detect air in-leakage that could dilute the samples.

Ammonia samples are analyzed with an ion-specific electrode. The instrument is calibrated using 0.1 µg/ml, 0.5 µg/ml, 1.0 µg/ml, 5.0 µg/ml and 10 µg/ml standards. At the conclusion of each analytical session, a sample spiked with a known amount of ammonia is analyzed and the result compared with the predicted concentration. A blank sample is analyzed to detect zero drift and a mid-range standard is also analyzed to detect calibration drift.

Both the sulfate (SO_2 & SO_3) samples and the chloride (HCl) samples are analyzed by means of ion chromatography. This instrument is calibrated using a minimum of three points on the calibration curve. As with the ammonia analysis, spiked samples, blanks, and standards are analyzed to ensure that calibration drift has not occurred.

While gas-phase flue gas constituents are being sampled, both field blanks and sample blanks are run as quality control checks. The field blank consists of a container of type 1 de-ionized water that is exposed to ambient air at the sampling site. Sample blanks are obtained by passing ambient air through the probe and through the impingers filled with the appropriate solution. These blanks would detect sample contamination should it occur.

Section 3

AIR HEATER TEST RESULTS

AMMONIA CONCENTRATIONS

Concentrations of ammonia were measured simultaneously at each of the three air heater inlet and outlet test locations with the reactors operating at Test Condition 22, the baseline operating condition (NH_3/NO_x = 0.8), and at Test Condition 24 (NH_3/NO_x = 1.0). The ammonia testing protocol, designed to differentiate between gas-phase and solid-phase ammonia, as described in Section 2, was used.

The test data and estimated solid-phase ammonia concentrations on a mass of ammonia per mass of ash basis are presented in Table 3-1. The mass-basis ammonia concentrations shown in the table for the ash entering and exiting the air heater were derived both from solid-phase ammonia measurements conducted simultaneously with the gas-phase measurements and mass concentration measurements made earlier at the same location. The ammonia partitioning between the gas and solid phases shown in the table are similar for all three air heaters. These data show that ammonia partitioning at the inlet to the air heater is roughly equal (on a flue gas volumetric basis or ammonia mass basis) between the gas and solid phases and that the ammonia shifts heavily to the solid phase as it cools while passing through the air heater.

As expected, the data in Table 3-1 show a significant increase in ammonia concentration in both the gas and solid phase (except the gas phase at the air heater outlets) during operation at the higher NH_3/NO_x ratio at Test Condition 24. For the gas phase at the air heater inlets, the range of concentrations at Test Condition 24 was 1.2 to 7.3 ppm(v) dry @ 3 % O_2, while at Test Condition 22 the range of concentrations was 0.4 to 1.3 ppm(v) dry @ 3 % O_2. For the solid phase at the air heater inlets the range of ammonia concentrations at Test Condition 24 was 1.4 to 6.5 ppm(v) dry @ 3 % O_2, (188 to 887 µg/g of ash), while at Test Condition 22 the range of ammonia concentrations was 0.4 to 2.0 ppm(v) dry @ 3 % O_2 (51 to 270 µg/g of ash). The partitioning shifts heavily toward the solid phase at the air heater outlets. As expected, a significant increase in total ammonia concentration was measured during operation at the higher NH_3/NO_x ratio at Test Condition 24.

Each ammonia sampling train was used to run a blank sample by sampling ambient air before each set of tests was run to demonstrate the integrity of the equipment. The original data summaries from these blank ammonia concentration tests can be found in Table A-1 in Appendix A. The original data summaries for the actual ammonia concentration tests at Test Conditions 22 and 24 can be found in Table A-2 in Appendix A.

SO₂ and SO₃ CONCENTRATIONS

Concentrations of sulfur dioxide and sulfur trioxide were measured simultaneously at the air heater inlet and outlet test locations of Reactors A, B, and C during operation at Test Condition 22. Tests were conducted both at the beginning (baseline performance) and the end of the second parametric test sequence under Task 4 (June and October 1994). Concentrations of sulfur dioxide and sulfur

Table 3-1. SCR Reactor Air Preheater Tests, September and October, 1994: Ammonia Concentrations

Reactor	Date	$\frac{NH_3}{NO_x}$	Air Preheater Location	Gas-Phase NH_3, ppm(v), @ 3% O_2, dry	Solid-Phase NH_3, Equivalent ppm(v), @ 3% O_2, dry	Estimated Solid-Phase NH_3, µg/g	Estimated Upper Limit Solid-Phase NH_3, µg/g	Estimated Lower Limit Solid-Phase NH_3, µg/g
A	13-Oct-94	0.8	Inlet	0.4 ± 0.1	0.7 ± 0.2	87	119	58
			Outlet	<0.6	1.4 ± 0.2	191	222	163
A	13-Oct-94	1.0	Inlet	1.2 ± 0.3	1.4 ± 0.2	188	226	152
			Outlet	<0.6	1.7 ± 0.1	233	252	215
B	12-Oct-94	0.8	Inlet	0.6 ± 0.1	0.4 ± 0.1	51	61	41
			Outlet	<0.5	0.8 ± 0.1	119	134	104
B	12-Oct-94	1.0	Inlet	7.3 ± 6.3	4.9 ± 3.7	692	1233	170
			Outlet	<0.5	6.5 ± 5.5	923	1731	141
C	29-Sep-94	0.8	Inlet	1.3 ± 0.4	2.0 ± 0.2	270	314	228
			Outlet	<0.5	2.4 ± 0.2	324	359	291
C	30-Sep-94	1.0	Inlet	3.6 ± 0.5	6.5 ± 0.8	887	1028	755
			Outlet	<0.5	7.2 ± 3.5	990	1523	491

trioxide were also measured in October 1994 at the split inlet upstream of the three large reactors. The test data are summarized in Table 3-2 (June 1994) and Table 3-3 (October 1994).

In June 1994 SO_3 concentrations at the air heater inlets ranged from 10 to 19 ppm(v) dry @ 3% O_2, while SO_3 concentrations at the air heater outlets ranged from 10 to 16 ppm(v) dry @ 3% O_2. Reductions in SO_3 concentration across the air heaters ranged from 0% on Reactor B to 11% on Reactor A to 32% on Reactor C. Within the tolerance of the standard deviations given for average SO_2 concentrations in Table 3-3, there was only a slight increase in SO_2 concentration across the individual air heaters. The reason for this slight increase in concentration is not known. Air heater inlet SO_2 concentrations averaged 2038 ± 191 ppm(v) dry @ 3% O_2, while the air heater outlet SO_2 concentrations averaged 2143 ± 179 ppm(v) dry @ 3% O_2. Table A-3 in Appendix A contains the original data summaries for these tests.

In October 1994 SO_2 and SO_3 concentrations were measured at the split inlet upstream of the three large reactors. The average SO_2 concentration was 1880 ± 11 ppm(v) dry @ 3% O_2 and the average SO_3 concentration was 1.3 ± 0.01 ppm(v) dry @ 3% O_2. Table A-4 in Appendix A contains the original data summaries for these tests.

During the October 1994 tests SO_3 concentrations at the air heater inlets ranged from 3.8 to 21.7 ppm(v) dry @ 3% O_2, while SO_3 concentrations at the air heater outlets ranged from 7.3 to 15.6 ppm(v) dry @ 3% O_2. Changes in SO_3 concentrations across the air heaters ranged from factors of 0.72 on Reactor A to 2.52 on Reactor B to 1.78 on Reactor C. Within the tolerance of the standard deviations given for average SO_2 concentrations in Table 3-3, there were measurable reductions in SO_2 concentrations across the individual air heaters. The reason for these rather large decreases in SO_2 concentration is not known. SO_2 concentrations changed by factors of 0.75 (Reactor A), 0.85 (Reactor B), and 0.76 (Reactor C). Overall, air heater inlet SO_2 concentrations averaged 1,957 ± 71 ppm(v) dry @ 3% O_2, while the air heater outlet SO_2 concentrations averaged 1,539 ± 165 ppm(v) dry @ 3% O_2. Table A-5 in Appendix A contains the original data summaries for these tests.

HCl CONCENTRATION

Measurements of the concentration of hydrogen chloride were conducted simultaneously at the inlet and outlet of each of the three large reactor air heaters in October 1994. Three independent tests were performed at each test location. Single point sampling was used. The test results are presented in Table 3-4 as the average concentration and the standard deviation. Air heater inlet HCl concentrations fell within the range of 92.8 to 101 ppm(v) dry @ 3% O_2. Air heater outlet HCl concentrations fell within the range of 81.1 to 93.1 ppm(v) dry @ 3% O_2. The decrease in HCl concentration across the air heaters ranged from 7.8% on Reactor B to 14.9% on Reactor C. The specific cause for this reduction in HCl concentration across the reactor air heaters is not known at this time. The original data summaries for these HCl tests are presented in Table A-6 in Appendix A.

PARTICULATE MASS CONCENTRATION

Particulate mass concentrations were measured only at the outlet of the air heaters on Reactors A and B during the baseline performance tests in June 1994. No outlet measurement was possible on

Table 3-2. SCR Reactor Air Heater Tests: SO₂ and SO₃ Concentrations (June 1994)

| Reactor | SO₃ | | SO₂ | |
| | Inlet | Outlet | Inlet | Outlet |
	ppm(v), dry @ 3% O₂	ppm(v), dry @ 3% O₂	ppm(v), dry @ 3% O₂	ppm(v), dry @ 3% O₂
A	18.3 ± 3.5	16.0 ± 1.4	2187 ± 55	2294 ± 24
B	10.0 ± 5.1	10.0 ± 1.4	2104 ± 458	2190 ± 27
C	18.8 ± 3.1	12.6 ± 0.7	1822 ± 33	1945 ± 24

Table 3-3. SCR Reactor Air Heater Tests: SO_2 and SO_3 Concentrations (October 1994)

| Reactor | SO_3 | | SO_2 | |
| | Inlet | Outlet | Inlet | Outlet |
	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2
A	21.7 ± 2.3	15.6 ± 3.1	1890 ± 67	1413 ± 48
B	3.8 ± 1.2	9.6 ± 1.8	2032 ± 31	1726 ± 92
C	4.1 ± 0.8	7.3 ± 0.5	1949 ± 14	1479 ± 46

Table 3-4. SCR Reactor Air Heater Tests: HCl Concentrations

Reactor	HCl	
	Inlet	Outlet
	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2
A	92.8 ± 1.9	81.1 ± 1.0
B	101 ± 5.0	93.1 ± 4.1
C	99.8 ± 4.0	84.9 ± 0.8

Table 3-5. SCR Reactor Air Heater Tests: Outlet Mass Concentrations

Reactor	Air Heater Outlet
	gr/dscf
A	2.48 ± 0.12
B	2.34 ± 0.04
C	Not Tested

the Reactor C air heater because sampling ports in a location suitable for mass concentration measurements could not be installed on that reactor.

Table 3-5 presents the mass concentration data for the air heater tests.   The average air heater outlet mass concentration was 2.48 ± 0.12 gr/dscf on Reactor A and 2.34 ± 0.04 gr/dscf on Reactor B. The original data summaries for the mass concentration tests can be found in Table A-7 in Appendix A.

Section 4

SUMMARY

This report is the eighth in a series of reports describing the results of testing and analytical services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The test results for the second sequence of air heater tests are summarized in this report. The air heater evaluations were conducted at the beginning (June 1994) and end (October 1994) of the second sequence of parametric testing under Task 4: Long-Term Parametric Tests. The tests in June 1994 (SO_2, SO_3, and particulate mass concentrations) were specifically conducted to establish a set of baseline performance values. (The air heaters had been washed and a partial replacement of baskets had occurred on the two rotary air heaters (A and B) during the May/June 1994 outage.)

The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The test facility includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 °F down to 300 °F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C incorporates a heat pipe design.

The air heater testing included the determinations of particulate mass concentration and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. Tests were performed at the inlet and outlet of each of the three air heaters (except for mass concentration measurements at the inlet of each reactor air heater and the outlet of the Reactor C air heater).

AMMONIA CONCENTRATIONS

At the air heater inlets, during measurements conducted in October 1994, the ammonia present was roughly divided equally between gas phase and solid phase. At the air heater outlets, the gas phase ammonia concentrations were below the detection limit, indicating that nearly all of the ammonia present had partitioned to the solid phase.

As expected, there was a measurable increase in ammonia concentration in both the gas and solid phases at the air heater inlets during operation at the higher NH_3/NO_x ratio at Test Condition 24 compared to Test Condition 22. At both parametric operating conditions, slip ammonia partitioned mostly to the solid phase at the air heater exit after being cooled while passing through the air heater.

SO$_2$ and SO$_3$ CONCENTRATIONS

In June 1994 (baseline preformance) SO$_3$ inlet concentrations ranged from 10 to 19 ppm(v) dry @ 3% O$_2$, while SO$_3$ concentrations at the reactor outlets ranged from 10 to 16 ppm(v) dry @ 3% O$_2$. Reductions in SO$_3$ concentration across the air heaters ranged from 0% on Reactor B to 11% on Reactor A to 32% on Reactor C. Within the tolerance of the standard deviations given for average SO$_2$ concentrations, there was only a slight increase in SO$_2$ concentration across the individual air heaters. The reason for this slight increase in concentration is not known. Air heater inlet SO$_2$ concentrations averaged 2038 ± 191 ppm(v) dry @ 3% O$_2$, while the air heater outlet SO$_2$ concentrations averaged 2143 ± 179 ppm(v) dry @ 3% O$_2$.

In October 1994 SO$_2$ and SO$_3$ concentrations were measured at the split inlet upstream of the three large reactors. The average SO$_2$ concentration was 1880 ± 11 ppm(v) dry @ 3% O$_2$ and the average SO$_3$ concentration was 1.3 ± 0.01 ppm(v) dry @ 3% O$_2$.

During the October 1994 tests SO$_3$ concentrations at the air heater inlets ranged from 3.8 to 21.7 ppm(v) dry @ 3% O$_2$, while SO$_3$ concentrations at the air heater outlets ranged from 7.3 to 15.6 ppm(v) dry @ 3% O$_2$. Changes in SO$_3$ concentrations across the air heaters ranged from factors of 0.72 on Reactor A to 2.52 on Reactor B to 1.78 on Reactor C. Within the tolerance of the standard deviations given for the average SO$_2$ concentrations, there were measurable reductions in SO$_2$ concentrations across the individual air heaters. The reason for these rather large decreases in SO$_2$ concentration is not known. SO$_2$ concentrations changed by factors of 0.75 (Reactor A), 0.85 (Reactor B), and 0.76 (Reactor C). Overall, air heater inlet SO$_2$ concentrations averaged 1,957 ± 71 ppm(v) dry @ 3% O$_2$, while the air heater outlet SO$_2$ concentrations averaged 1,539 ± 165 ppm(v) dry @ 3% O$_2$.

HCl CONCENTRATIONS

Air heater inlet HCl concentrations, measured in October 1994, fell within the range of 92.8 to 101 ppm(v) dry @ 3% O$_2$. Air heater outlet HCl concentrations fell within the range of 81.1 to 93.1 ppm(v) dry @ 3% O$_2$. The decrease in HCl concentration across the air heaters ranged from 7.8% on Reactor B to 14.9% on Reactor C. The specific cause for this reduction in HCl concentration across the reactor air heaters is not known at this time.

MASS CONCENTRATIONS

The average air heater outlet mass concentrations, measured during baseline performance tests in June 1994, were 2.48 ± 0.12 gr/dscf for the Reactor A air heater and 2.34 ± 0.04 gr/dscf for the Reactor B air heater.

APPENDIX A

ORIGINAL DATA SUMMARIES

Table A-1

	Time, Begin	Time, End	Date	NH₃ Concentration ppm(v), dry Average	σ	Analytical NH₃ mg/ml	Dil	Vol. ml	NH₃ µM	NH₃ ppm, dry	Dry Gas Meter Volume, Total STDI, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
093 AAI NHx 01	0950	1003	13-Oct	0.48	<	0.014 / 0.010	1	234	1.67 / 1.67	0.5 / 0.5	78.05	297	29.98	75	76	631.700	634.700	Stack 1	0.998
093 AAO NHx 01	0933	0944	13-Oct	0.75	<	0.000 / 0.000	1	366	2.61 / 2.61	0.8 / 0.8	77.86	295	29.98	72	72	530.200	533.200	SRI 3	0.989
093 BAI NHX 01	0926	0941	12-Oct	0.47	<	0.032 / 0.043	1	232	1.66 / 1.66	0.5 / 0.5	78.41	296	29.95	72	73	589.700	592.700	Stack 1	0.998
093 BAO NHx 01	0923	0934	12-Oct	0.54	<	0.010 / 0.009	1	301	2.15 / 2.15	0.5 / 0.5	89.08	292	29.95	66	67	485.900	489.300	SRI 3	0.989
091 CAI NHx 01	0939	0955	29-Sep	0.37	<	0.022 / 0.016	1	238	1.70 / 1.70	0.4 / 0.4	103.33	301	30.04	82	82	435.100	439.100	Stack 1	1.001
091 CAO NHx 01	0957	1012	29-Sep	0.60	<	0.073 / 0.076	1	287	2.05 / 2.05	0.6 / 0.6	76.86	300	30.04	80	80	387.800	390.800	SRI 3	0.989
091 CAO NHX 05	0859	0911	30-Sep	0.64	<	0.009 / 0.007	1	311	2.22 / 2.22	0.6 / 0.6	77.65	297	30.04	73	76	410.700	413.700	SRI 3	0.989
091 CAI NHx 05	0856	0612	30-Sep	0.30	<	0.028 / 0.022	1	191	1.36 / 1.36	0.3 / 0.3	103.53	300	30.04	81	81	460.200	464.200	Stack 1	1.001

Table A-2 (Reference Table 3-1)

	Time, Begin	Time, End	Date	NH₃ Concentration ppm(v), dry @3% O₂ Average	σ	Analytical NH₃ µg/ml	Dil	Vol ml	NH₃ µM	NH₃ ppm(v) dry, 3% O₂	Condition Number	Flow scfm	Temp °F	NH₃/NO₄	UnM5 Boiler Load MW	Sample O₂ %	STD Vol Total, dry	Temp Avg °K	Barometric Pressure In.Hg	Temp Begin °F	Temp End °F	Volume Begin ft³	Volume End ft³	Meter ID	Corr. Factor
093 AAI NHG 02	1043	1120	13-Oct-94	0.3 <		0.089 / 0.097	1	334	2.4 / 2.4	0.3 / 0.3	22	5000	705	0.8	79	2.8	156.25	297	29.98	77	73	635.000	641.000	Stack 1	0.998
093 AAI NHG 03	1135	1215	13-Oct-94	0.5 ± 0.004		0.179 / 0.177		304	3.9 / 3.8	0.5 / 0.5	22	5000	705	0.8	79	2.6	157.13	295	29.98	73	71	641.200	647.200	Stack 1	0.998
093 AAI NHG 04	1237	1316	13-Oct-94	0.4 ± 0.004		0.118 / 0.116		305	2.6 / 2.5	0.4 / 0.4	22	5000	705	0.8	76	2.6	156.98	296	29.98	72	73	647.400	653.400	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @ 3% O₂				0.4 ± 0.11							Testing Interval Averages / σ				78 / 1.7	2.7 / 0.1									
093 AAI NHS 02	1043	1120	13-Oct-94	0.4 ± 0.02		0.196 / 0.208	1	208	2.9 / 3.1	0.4 / 0.4	22	5000	705	0.8	79	2.8	156.25	297	29.98	77	73	635.000	641.000	Stack 1	0.998
093 AAI NHS 03	1135	1215	13-Oct-94	0.8 ± 0.005		0.363 / 0.366		215	5.6 / 5.6	0.8 / 0.8	22	5000	705	0.8	79	2.6	157.13	295	29.98	73	71	641.200	647.200	Stack 1	0.998
093 AAI NHS 04	1237	1316	13-Oct-94	0.8 ± 0.01		0.348 / 0.353		216	5.4 / 5.4	0.7 / 0.8	22	5000	705	0.8	76	2.6	156.98	296	29.98	72	73	647.400	653.400	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @ 3% O₂				0.7 ± 0.20							Testing Interval Averages / σ				78 / 1.7	2.7 / 0.1									
093 AAO NHG 02	1043	1112	13-Oct-94	0.6 <		0.000 / 0.000	1	380	2.7 / 2.7	0.6 / 0.6	22	5000	705	0.8	79	8.8	156.02	295	29.98	70	72	534.200	540.200	SRI 3	0.9891
093 AAO NHG 03	1135	1204	13-Oct-94	0.8 <		0.000 / 0.000		372	2.7 / 2.7	0.6 / 0.6	22	5000	705	0.8	79	9.2	155.73	295	29.98	73	71	540.500	546.500	SRI 3	0.9891
093 AAO NHG 04	1236	1304	13-Oct-94	0.6 <		0.000 / 0.000		380	2.7 / 2.7	0.6 / 0.6	22	5000	705	0.8	76	8.6	156.91	293	29.98	68	68	546.800	552.800	SRI 3	0.9891
Average NH₃ Concentration, ppm(v), dry @ 3% O₂				0.6							Testing Interval Averages / σ				78 / 1.7	8.9 / 0.3									
093 AAO NHS 02	1043	1112	13-Oct-94	1.6 ± 0.02		0.502 / 0.512	1	208	7.5 / 7.6	1.6 / 1.6	22	5000	705	0.8	79	8.8	156.02	295	29.98	70	72	534.200	540.200	SRI 3	0.9891
093 AAO NHS 03	1135	1204	13-Oct-94	1.4 ± 0.03		0.457 / 0.470		192	6.3 / 6.4	1.4 / 1.4	22	5000	705	0.8	79	9.2	155.73	295	29.98	73	71	540.500	546.500	SRI 3	0.9891
093 AAO NHS 04	1236	1304	13-Oct-94	1.3 ± 0.03		0.397 / 0.386		224	6.4 / 6.2	1.3 / 1.3	22	5000	705	0.8	76	8.6	156.91	293	29.98	68	68	546.800	552.800	SRI 3	0.9891
Average NH₃ Concentration, ppm(v), dry @ 3% O₂				1.4 ± 0.15							Testing Interval Averages / σ				78 / 1.7	8.9 / 0.3									

A-3

Table A-2, continued (Reference Table 3-1)

	Time, Begin	Time, End	Date	NH₃ Concentration, ppm(v), dry @ 3% O₂ Average		σ	Analytical NH₃ µg/ml	Dil	Vol ml	NH₃ µM	NH₃ ppm(v), dry, 3% O₂	Reactor Conditions Condition Number	Flow acfm	Temp °F	NH₃/NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Dry Gas Meter Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D.	Corr. Factor
093 AAI NHG 05	1441	1519	13-Oct-94	1.5	±	0.01	0.573 0.577	1	261	10.7 10.8	1.5 1.5	24	400	705	1	77	2.9	155.52	298	29.98	76	79	653.800	659.800	Stack 1	0.998
093 AAI NHG 06	1534	1613	13-Oct-94	1.1	±	0.01	0.439 0.438		244	7.7 7.6	1.1 1.1	24	400	705	1	77	2.8	154.66	300	29.98	80	81	660.000	666.000	Stack 1	0.998
093 AAI NHG 07	1626	1704	13-Oct-94	1.0	±	0.80	0.129 0.450		346	3.2 11.1	0.5 1.6	24	400	705	1	78	2.6	154.51	300	29.98	81	81	666.200	672.200	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @ 3% O2				1.2	±	0.28						Testing Interval Averages / σ				77 0.6	2.8 0.2									
093 AAI NHS 05	1441	1519	13-Oct-94	1.2	±	0.004	0.443 0.441	1	276	8.7 8.7	1.3 1.2	24	400	705	1	77	2.9	155.52	298	29.98	76	79	653.800	659.800	Stack 1	0.998
093 AAI NHS 06	1534	1613	13-Oct-94	1.3	±	0.03	0.584 0.605		217	9.1 9.4	1.3 1.3	24	400	705	1	77	2.8	154.66	300	29.98	80	81	660.000	666.000	Stack 1	0.998
093 AAI NHS 07	1626	1704	13-Oct-94	1.6	±	0.00	0.717 0.719		226	11.6 11.6	1.6 1.6	24	400	705	1	78	2.6	154.51	300	29.98	81	81	666.200	672.200	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @ 3% O3				1.4	±	0.21						Testing Interval Averages / σ				77 0.6	2.8 0.2									
093 AAO NHG 05	1439	1512	13-Oct-94	< 0.7			0.000 0.000	1	393	2.8 2.8	0.7 0.7	24	400	705	1	77	10.1	154.42	298	29.98	75	78	553.700	559.700	SRI 3	0.9891
093 AAO NHG 06	1535	1605	13-Oct-94	< 0.6			0.000 0.000		403	2.9 2.9	0.6 0.6	24	400	705	1	77	9.3	153.70	299	29.98	78	80	560.000	566.000	SRI 3	0.9891
093 AAO NHG 07	1625	1652	13-Oct-94	< 0.5			0.000 0.000		315	2.3 2.3	0.5 0.5	24	400	705	1	78	9.0	153.70	299	29.98	78	80	566.300	572.300	SRI 3	0.9891
Average NH₃ Concentration, ppm(v), dry @ 3% O4				< 0.6								Testing Interval Averages / σ				77 0.6	9.5 0.6									
093 AAO NHS 05	1439	1512	13-Oct-94	1.8	±	0.02	0.536 0.543	1	194	7.4 7.5	1.8 1.8	24	400	705	1	77	10.1	154.42	298	29.98	75	78	553.700	559.700	SRI 3	0.9891
093 AAO NHS 06	1535	1605	13-Oct-94	1.7	±	0.02	0.455 0.464		228	7.4 7.6	1.7 1.7	24	400	705	1	77	9.3	153.70	299	29.98	78	80	560.000	566.000	SRI 3	0.9891
093 AAO NHS 07	1625	1652	13-Oct-94	1.7	±	0.01	0.534 0.528		210	8.0 7.9	1.8 1.7	24	400	705	1	78	9.0	153.70	299	29.98	78	80	566.300	572.300	SRI 3	0.9891
Average NH₃ Concentration, ppm(v), dry @ 3% O5				1.7	±	0.06						Testing Interval Averages / σ				77 0.6	9.5 0.6									

A-4

Table A-2, continued (Reference Table 3-1)

	Time, Begin	Time, End	Date	NH₃ Conc ppm(v), dry @3% O₂ Average	±	σ	NH₃ mg/ml	Dil	Vol. ml	NH₃ µM	NH₃ ppm(v), dry 3% O₂	Condition Number	Flow acfm	Temp, °F	NH₃/NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Volume Total STD L, dry	Temp, Avg. °K	Barometric Pressure in.Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D.	Corr. Factor
093 BAI NHG 02	1043 Possible leak^A	1118	12-Oct	0.8	±	0.03	0.187	1	277	3.7	0.6	22	5000	705	0.8	84	4.0	156.09	297	29.95	74	76	593.000	599.000	Stack 1	0.998
							0.200			4.0	0.6															
093 BAI NHG 03	1137	1215	12-Oct	0.6	±	0.00	0.182		302	3.9	0.6	22	5000	705	0.8	86	4.2	157.95	298	29.95	76	79	599.200	605.300	Stack 1	0.998
							0.180			3.9	0.6															
093 BAI NHG 04	1233	1312	12-Oct	0.7	±	0.01	0.204		313	4.6	0.7	22	5000	705	0.8	83	4.2	154.93	299	29.95	78	80	605.500	611.500	Stack 1	0.998
							0.199			4.4	0.7															
Average NH₃ Concentration, ppm(v), dry @3% O2				0.6	±	0.06						Testing Interval Averages				84 / σ 1.5	4.1 / σ 0.1									
093 BAI NHS 02	1043 Possible leak^A	1118	12-Oct	0.3	±	0.00	0.145	1	181	1.9	0.3	22	5000	705	0.8	84	4.0	156.09	297	29.95	74	76	593.000	599.000	Stack 1	0.998
							0.147			1.9	0.3															
093 BAI NHS 03	1137	1215	12-Oct	0.4	±	0.01	0.184		193	2.5	0.4	22	5000	705	0.8	88	4.2	157.95	298	29.85	76	79	599.200	605.300	Stack 1	0.998
							0.191			2.6	0.4															
093 BAI NHS 04	1233	1312	12-Oct	0.4	±	0.00	0.171		213	2.6	0.4	22	5000	705	0.8	83	4.2	154.93	299	29.95	78	80	605.500	611.500	Stack 1	0.998
							0.171			2.6	0.4															
Average NH₃ Concentration, ppm(v), dry @3% O2				0.4	±	0.06						Testing Interval Averages				84 / σ 1.5	4.1 / σ 0.1									
093 BAO NHG 02	1042	1111	12-Oct	< 0.5			0.030	1	323	2.3	0.5	22	5000	705	0.8	84	8.0	156.31	294	29.95	68	71	489.700	495.700	SRI 3	0.9891
							0.027			2.3	0.5															
093 BAO NHG 03	1137	1206	12-Oct	< 0.5			0.016		354	2.5	0.5	22	5000	705	0.8	86	7.5	156.60	293	29.95	67	70	496.200	502.200	SRI 3	0.9891
							0.017			2.5	0.5															
093 BAO NHG 04	1234	1303	12-Oct	< 0.5			0.014		371	2.7	0.5	22	5000	705	0.8	83	7.3	156.01	294	29.95	69	72	502.700	508.700	SRI 3	0.9891
							0.014			2.7	0.5															
Average NH₃ Concentration, ppm(v), dry @3% O2				< 0.5								Testing Interval Averages				84 / σ 1.5	7.6 / σ 0.4									
093 BAO NHS 02	1042	1111	12-Oct	0.7	±	0.01	0.348	1	148	3.7	0.7	22	5000	705	0.8	84	8.0	156.31	294	29.95	68	71	489.700	495.700	SRI 3	0.9891
							0.355			3.8	0.7															
093 BAO NHS 03	1137	1206	12-Oct	0.9	±	0.01	0.426		158	4.8	0.9	22	5000	705	0.8	86	7.5	156.60	293	29.85	67	70	496.200	502.200	SRI 3	0.9891
							0.435			4.9	0.9															
093 BAO NHS 04	1234	1303	12-Oct	0.8	±	0.01	0.344		184	4.5	0.9	22	5000	705	0.8	83	7.3	156.01	294	29.95	69	72	502.700	508.700	SRI 3	0.9891
							0.340			4.5	0.8															
Average NH₃ Concentration, ppm(v), dry @3% O2				0.8	±	0.10						Testing Interval Averages				84 / σ 1.5	7.6 / σ 0.4									

Table A-2, continued (Reference Table 3-1)

	Time, Begin	Time, End	Date	NH₃ Conc. ppm(v), dry, @3% O₂ Average	σ	Analytical NH₃ mg/ml	Dil	Vol. ml	NH₃ µM	NH₃ ppm(v), dry, 3% O₂	Condition Number	Flow scfm	Temp, °F	NH₃/NO₂	Unit 5 Boiler Load, MW	Sample O₂ %	Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D.	Corr. Factor
093 BAJ NHG 05	1457	1535	12-Oct	2.2	± 0.03	0.660 / 0.673	1	299	14.1 / 14.4	2.2 / 2.2	24	400	705	1	82	4.4	155.65	298	29.95	77	76	611.700	617.700	Stack 1	0.998
093 BAJ NHG 06	1554	1633	12-Oct	5.5	± 0.05	1.540 / 1.560		316	34.8 / 35.2	5.4 / 5.5	24	400	705	1	80	4.4	155.95	297	29.95	76	75	617.900	623.900	Stack 1	0.998
093 BAJ NHG 07	1648	1727	12-Oct	14.3	± 0.00	4.790 / 4.790		272	93.1 / 93.1	14.3 / 14.3	24	400	705	1	80	4.2	155.80	297	29.95	76	76	624.100	630.100	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @3% O2				7.3	± 6.28						Testing Interval Averages / σ				81 / 1.2	4 / 0.1									
093 BAJ NHS 05	1457	1535	12-Oct	1.8	± 0.07	0.809 / 0.763	1	204	11.8 / 11.1	1.8 / 1.7	24	400	705	1	82	4.4	155.65	298	29.95	77	76	611.700	617.700	Stack 1	0.998
093 BAJ NHS 06	1554	1633	12-Oct	3.9	± 0.00	1.770 / 1.770		200	25.3 / 25.3	3.9 / 3.9	24	400	705	1	80	4.4	155.95	297	29.95	76	75	617.900	623.900	Stack 1	0.998
093 BAJ NHS 07	1648	1727	12-Oct	8.9	± 0.05	3.830 / 3.800		213	58.3 / 57.8	9.0 / 8.9	24	400	705	1	80	4.2	155.80	297	29.95	76	76	624.100	630.100	Stack 1	0.998
Average NH₃ Concentration, ppm(v), dry @3% O3				4.9	± 3.67						Testing Interval Averages / σ				81 / 1.2	4 / 0.1									
093 BAO NHG 05	1456	1528	12-Oct	0.5	< 0.02	0.008 / 0.008	1	355	2.5 / 2.5	0.5 / 0.5	24	400	705	1	82	7.5	154.41	297	29.95	74	78	509.400	515.400	SRI 1	0.9891
093 BAO NHG 06	1556	1624	12-Oct	0.5	< 0.03	0.009 / 0.009		351	2.5 / 2.5	0.5 / 0.5	24	400	705	1	80	7.3	153.84	299	29.95	76	80	516.000	522.000	SRI 1	0.9891
093 BAO NHG 07	1646	1720	12-Oct	0.5	< 0.03	0.010 / 0.049		342	2.4 / 2.4	0.5 / 0.5	24	400 ·	705	1	80	7.0	154.12	298	29.95	78	76	522.500	528.500	SRI 1	0.9891
Average NH₃ Concentration, ppm(v), dry @3% O4				0.5	<						Testing Interval Averages / σ				81 / 1.2	7.3 / 0.3									
093 BAO NHS 05	1456	1528	12-Oct	2.3	± 0.02	0.869 / 0.878	1	191	11.9 / 12.0	2.3 / 2.3	24	400	705	1	82	7.5	154.41	297	29.95	74	78	509.400	515.400	SRI 1	0.9891
093 BAO NHS 06	1556	1624	12-Oct	4.5	± 0.03	2.820 / 2.790		117	23.6 / 23.3	4.5 / 4.5	24	400	705	1	80	7.3	153.84	299	29.95	76	80	516.000	522.000	SRI 1	0.9891
093 BAO NHS 07	1646	1720	12-Oct	12.7	± 0.03	5.230 / 5.250		182	68.0 / 68.3	12.7 / 12.8	24	400	705	1	80	7.0	154.12	298	29.95	78	76	522.500	528.500	SRI 1	0.9891
Average NH₃ Concentration, ppm(v), dry @3% O5				6.5	± 5.51						Testing Interval Averages / σ				81 / 1.2	7.3 / 0.3									

Table A-2, continued (Reference Table 3-1)

ID	Time Begin	Time End	Date	NH₃ Conc. Average ppm(v), dry @ 3% O₂	±	σ	NH₃ mg/ml	Dil	Vol ml	NH₃ µM	NH₃ ppm(v), dry, 3% O₂	Condition Number	Flow acfm	Temp °F	NH₃/NOx	Unit 5 Boiler Load, MW	Sample O₂ %	Volume Total STD, dry	Temp Avg °K	Barometric Pressure in.Hg	Temp Begin °F	Temp End °F	Volume Begin ft³	Volume End ft³	Meter ID	Corr. Factor
091 CAJ NHG 02	1028	1104	29-Sep	1.7	±	0.03	0.311 / 0.318	1	368	8.2 / 8.4	1.7 / 1.7	22	5000	705	0.8	85	2.0	102.76	302	30.04	83	87	439.300	443.300	Stack 1	1.001
091 CAJ NHG 03	1259	1335	29-Sep	1.4	±	0.00	0.516 / 0.518		255	9.4 / 9.4	1.4 / 1.4	22	5000	705	0.8	85	2.6	152.47	306	30.04	90	92	445.600	451.600	Stack 1	1.001
091 CAJ NHG 04	1354	1432	29-Sep	0.8	±	0.01	0.280 / 0.285		288	5.8 / 5.9	0.8 / 0.8	22	5000	705	0.8	85	2.5	151.78	307	30.04	93	94	451.800	457.800	Stack 1	1.001
Average NH₃ Concentration, ppm(v), dry @ 3% O2				1.3	±	0.44						Testing Interval Averages				85 / σ	2.0 / σ									
091 CAJ NHS 02	1028	1104	29-Sep	2.1	±	0.05	0.650 / 0.671	1	218	10.1 / 10.4	2.1 / 2.2	22	5000	705	0.8	85	2.0	102.76	302	30.04	83	87	439.300	443.300	Stack 1	1.001
091 CAJ NHS 03	1259	1335	29-Sep	1.7	±	0.00	0.781 / 0.781		209	11.7 / 11.7	1.7 / 1.7	22	5000	705	0.8	85	2.6	152.47	306	30.04	90	92	445.600	451.600	Stack 1	1.001
091 CAJ NHS 04	1354	1432	29-Sep	2.1	±	0.02	0.953 / 0.969		212	14.4 / 14.7	2.1 / 2.1	22	5000	705	0.8	85	2.5	151.78	307	30.04	93	94	451.800	457.800	Stack 1	1.001
Average NH₃ Concentration, ppm(v), dry @ 3% O2				2.0	±	0.25						Testing Interval Averages				85 / σ	2.0 / σ									
091 CAO NHG 02	1033	1104	29-Sep	0.5	<	0.01	0.033 / 0.031	1	388	2.8 / 2.8	0.5 / 0.5	22	5000	705	0.8	85	5.0	152.59	302	30.04	82	86	391.200	397.200	SRI 3	0.989
091 CAO NHG 03	1304	1335	29-Sep	0.5	<	0.03	0.027 / 0.025		420	3.0 / 3.0	0.5 / 0.5	22	5000	705	0.8	85	5.3	151.20	305	30.04	87	91	397.500	403.500	SRI 3	0.989
091 CAO NHG 04	1359	1429	29-Sep	0.4			0.014 / 0.011		338	2.4 / 2.4	0.4 / 0.4	22	5000	705	0.8	85	5.2	150.24	307	30.04	92	93	404.000	410.000	SRI 3	0.989
Average NH₃ Concentration, ppm(v), dry @ 3% O2				0.5	<							Testing Interval Averages				85 / σ	5.0 / σ									
091 CAO NHS 02	1033	1104	29-Sep	2.2	±	0.01	0.736 / 0.742	1	248	13.0 / 13.1	2.2 / 2.2	22	5000	705	0.8	85	5.0	152.59	302	30.04	82	86	391.200	397.200	SRI 3	0.989
091 CAO NHS 03	1304	1335	29-Sep	2.4	±	0.03	0.775 / 0.788		257	14.2 / 14.5	2.4 / 2.4	22	5000	705	0.8	85	5.3	151.20	305	30.04	87	91	397.500	403.500	SRI 3	0.989
091 CAO NHS 04	1359	1429	29-Sep	2.5	±	0.03	0.843 / 0.856		240	14.5 / 14.7	2.5 / 2.5	22	5000	705	0.8	85	5.2	150.24	307	30.04	92	93	404.000	410.000	SRI 3	0.989
Average NH₃ Concentration, ppm(v), dry @ 3% O2				2.4	±	0.17						Testing Interval Averages				85 / σ	5.0 / σ									

Table A-2, continued (Reference Table 3-1)

Sample	Time, Begin	Time, End	Date	NH₃ Conc. Average (ppm(v), dry @3% O₂)	± σ	Analytical NH₃ mg/ml	Dil	Vol. ml	NH₃ µM	NH₃ ppm(v), dry, 3% O₂	Condition Number	Flow scfm	Temp °F	NH₃/NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Volume Total STD l, dry	Temp Avg °K	Barometric Pressure in.Hg	Temp Begin °F	Temp End °F	Volume Begin ft³	Volume End ft³	Meter ID	Corr. Factor
091 CAJ NHG 06	0921	0957	30-Sep	3.1	± 0.07	1.21 / 1.25	1	257	22.2 / 22.9	3.1 / 3.2	24	400	705	1	86	2.3	154.86	301	30.04	80	85	464.600	470.600	Stack 1	1.001
091 CAJ NHG 07	1020	1058	30-Sep	3.7	± 0.02	1.27 / 1.28		283	25.7 / 25.9	3.6 / 3.7	24	400	705	1	86	2.5	153.44	304	30.04	86	89	470.800	476.800	Stack 1	1.001
091 CAJ NHG 08	1239	1315	30-Sep	4.1	± 0.01	2.43 / 2.42		254	44.1 / 43.9	4.1 / 4.1	24	400	705	1	87	1.9	226.23	309	30.04	96	98	477.000	486.000	Stack 1	1.001
Testing Interval Averages σ															86 / 1	2 / 0									
Average NH₃ Concentration, ppm(v), dry @3% O2				3.6	± 0.48																				
091 CAJ NHS 06	0921	0957	30-Sep	6.5	± 0.00	3.12 / 3.12	1	208	46.4 / 46.4	6.5 / 6.5	24	400	705	1	86	2.3	154.86	301	30.04	80	85	464.600	470.600	Stack 1	1.001
091 CAJ NHS 07	1020	1058	30-Sep	7.2	± 0.01	3.44 / 3.45		207	50.9 / 51.0	7.2 / 7.2	24	400	705	1	86	2.5	153.44	304	30.04	86	89	470.800	476.800	Stack 1	1.001
091 CAJ NHS 08	1239	1315	30-Sep	5.7	± 0.08	4.21 / 4.13		204	61.3 / 60.2	5.7 / 5.6	24	400	705	1	87	1.9	226.23	309	30.04	96	98	477.000	486.000	Stack 1	1.001
Testing Interval Averages σ															86 / 1	2 / 0									
Average NH₃ Concentration, ppm(v), dry @3% O3				6.5	± 0.78																				
091 CAO NHG 06	0926	0957	30-Sep	0.5	✓ 0.00	0.004 / 0.008	1	377	2.7 / 2.7	0.5 / 0.5	24	400	705	1	86	5.7	153.44	300	30.04	78	84	414.300	420.300	SRI 3	0.989
091 CAO NHG 07	1025	1053	30-Sep	0.5	± 0.05	0.094 / 0.116		333	2.4 / 2.8	0.4 / 0.5	24	400	705	1	86	6.9	152.17	303	30.04	85	86	420.700	426.700	SRI 3	0.989
091 CAO NHG 08	1244	1316	30-Sep	0.5	✓	0.086 / 0.080		388	2.8 / 2.8	0.5 / 0.5	24	400	705	1	87	6.7	150.38	306	30.04	90	94	427.200	433.200	SRI 3	0.989
Testing Interval Averages σ															86 / 1	6 / 1									
Average NH₃ Concentration, ppm(v), dry @3% O4				0.5	✓																				
091 CAO NHS 06	0926	0957	30-Sep	4.4	± 0.00	1.54 / 1.54	1	232	25.5 / 25.5	4.4 / 4.4	24	400	705	1	86	5.7	153.44	300	30.04	78	84	414.300	420.300	SRI 3	0.989
091 CAO NHS 07	1025	1053	30-Sep	11.1	± 0.00	3.65 / 3.65		227	59.2 / 59.2	11.1 / 11.1	24	400	705	1	86	6.9	152.17	303	30.04	85	86	420.700	426.700	SRI 3	0.989
091 CAO NHS 08	1244	1316	30-Sep	6.1	± 0.06	2.63 / 2.67		171	32.1 / 32.6	6.0 / 6.1	24	400	705	1	87	6.7	150.38	306	30.04	90	94	427.200	433.200	SRI 3	0.989
Testing Interval Averages σ															86 / 1	6 / 1									
Average NH₃ Concentration, ppm(v), dry @3% O5				7.2	± 3.51																				

Table A-3

Dry Gas Meter / Analytical / Concentration

Week Site	Time, Begin	Time, End	Date	Concentration ppm(v), dry, @3% O2 Average	σ	SO₄⁼ mg/ml	Dil	Vol. ml	SO₄⁼ μM	SO₂, or SO₃ ppm, dry, @3% O2	Unit 5 Boiler Load, MW	Sample O₂ %	Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In.Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
092 SI SO₃ 01	1004	1018	7-Oct-94	1.1	± 0.00	4.199 / 4.210	1	95.4	4 / 4	1.1 / 1.1	88	1.8	77.57	301	30.13	79	84	536.100	539.100	Stack 1	0.998
092 SI SO₃ 02	1030	1043	7-Oct-94	1.3	± 0.00	3.868 / 3.888	1	114	5 / 5	1.3 / 1.3	88	1.8	76.58	304	30.13	86	91	539.300	542.300	Stack 1	0.998
092 SI SO₃ 03	1054	1107	7-Oct-94	1.3	± 0.00	4.730 / 4.734	1	93.6	5 / 5	1.3 / 1.3	89	1.8	76.23	306	30.13	89	93	542.600	545.600	Stack 1	0.998
Average SO₃ Concentration, ppm(v), dry @3% O2				1.3	± 0.01						Average 88, σ 1	1.8 / 0.0									

Week Site	Time, Begin	Time, End	Date	Concentration ppm(v), dry, @3% O2 Average	σ	SO₄⁼ mg/ml	Dil	Vol. ml	SO₄⁼ μM	SO₂, or SO₃ ppm, dry, @3% O2
092 SI SO₂ 01	1004	1018	7-Oct-94	1876	± 9.8	4.898 / 4.862	200	682	6959 / 6908	1883 / 1869
092 SI SO₂ 02	1030	1043	7-Oct-94	1893	± 3.3	4.498 / 4.509		736	6897 / 6914	1891 / 1895
092 SI SO₂ 03	1054	1107	7-Oct-94	1872	± 1.9	4.794 / 4.801		680	6792 / 6801	1870 / 1873
Average SO₂ Concentration, ppm(v), dry @3% O2				1880	± 11.2					

A-9

Table A-4 (Reference Table 3-2)

076 AAI SO3

	Time, Begin	Time, End	Date, 1994	Concentration ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil ml	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm, dry @3% O2	Condition Number	Flow scfm	Temp °F	Nb3 NO1	Unit 5 Boiler Load MW	Sample O2 %	Volume, Total	STD I, dry	Temp, Avg. °K	Barometric Pressure in. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
076 AAI SO3 01	1320	1335	15-Jun-94	14.3 ±	0.10	1.648 / 1.664	25	109	47 / 47	14.3 / 14.4	22	5	700	0.8	84	3.4	75.05		312	30.17	101	102	614.300	617.300	Stack 1	1.0000
076 AAI SO3 02	1349	1406	15-Jun-94	20.1 ±	0.19	1.888 / 1.981		128	68 / 65	20.2 / 20.0	22	5	700	0.8	84	3.4	74.99		312	30.17	102	102	617.700	620.700	Stack 1	1.0000
076 AAI SO3 03	1427	1441	15-Jun-94	20.5 ±	0.23	2.438 / 2.477		108	69 / 70	20.4 / 20.7	22	5	700	0.8	84	2.9	74.99		312	30.17	103	101	321.300	324.300	Stack 1	1.0000

Average SO3 Concentration, ppm(v), wet @ 3% O2: 20.3 ± 0.3

Testing Interval Averages: MW 84, σ 0; O2 3.2, σ 0.3

076 AAI SO2

	Time, Begin	Time, End	Date, 1994	Concentration ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil ml	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm, dry @3% O2
076 AAI SO2 01	1320	1335	15-Jun-94	2191.6 ±	0.64	4.814 / 4.812	200	716	7181 / 7178	2192.1 / 2191.2
076 AAI SO2 02	1349	1406	15-Jun-94	2238.8 ±	4.71	5.386 / 5.370		654	7338 / 7317	2242.2 / 2235.5
076 AAI SO2 03	1427	1441	15-Jun-94	2129.4 ±	22.83	4.588 / 4.517		758	7223 / 7114	2145.6 / 2113.3

Average SO2 Concentration, ppm(v), wet @ 3% O2: 2184 ± 77

076 AAO SO3

	Time, Begin	Time, End	Date, 1994	Concentration ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil ml	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm, dry @3% O2	Condition Number	Flow scfm	Temp °F	Nb3 NO1	Unit 5 Boiler Load MW	Sample O2 %	Volume, Total	Temp, Avg. °K	Barometric Pressure in. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
076 AAO SO3 01	1325	1338	15-Jun-94	14.5 ±	0.04	1.633 / 1.639	25	83.8	38 / 38	14.5 / 14.5	22	5	700	0.8	84	7.5	73.69	314	30.17	106	106	58.600	61.600	Stack 2	0.9897
076 AAO SO3 02	1408	1420	15-Jun-94	17.1 ±	0.04	2.152 / 2.144		74.8	42 / 42	17.1 / 17.0	22	5	700	0.8	84	7.6	73.89	313	30.17	105	104	61.900	64.900	Stack 2	0.9897
076 AAO SO3 03	1434	1447	15-Jun-94	18.5 ±	0.03	1.917 / 1.922		81.7 / 41	41 / 41	18.5 / 18.6	22	5	700	0.8	84	7.5	73.89	313	30.17	104	105	65.200	68.200	Stack 2	0.9897

Average SO3 Concentration, ppm(v), wet @ 3% O2: 16.8 ± 0.4

Testing Interval Averages: MW 84, σ 0; O2 7.5, σ 0.1

076 AAO SO2

	Time, Begin	Time, End	Date, 1994	Concentration ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil ml	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm, dry @3% O2
076 AAO SO2 01	1325	1338	15-Jun-94	2269.5 ±	26.65	4.479 / 4.554	200	594	5543 / 5636	2250.6 / 2288.3
076 AAO SO2 02	1408	1420	15-Jun-94	2294.3 ±	2.89	4.487 / 4.495		601	5618 / 5628	2292.3 / 2296.4
076 AAO SO2 03	1434	1447	15-Jun-94	2317.4 ±	2.68	4.275 / 4.282		642	5718 / 5727	2315.6 / 2319.3

Average SO2 Concentration, ppm(v), wet @ 3% O2: 2306 ± 16

Table A-4, continued (Reference Table 3-2)

BAI SO3

Sample	Time Begin	Time End	Date 1994	ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm dry, @3% O2	Condition Number	Flow scfm	Temp °F	NH3 NO	Unit 5 Boiler Load, MW	Sample O2 %	Volume Total STD l, dry	Temp Avg °K	Barometric Pressure In. Hg	Temp Begin °F	Temp End °F	Volume Begin ft³	Volume End ft³	Meter I.D	Corr. Factor
076 BAI SO3 01	1307	1320	14-Jun-94	5.9	± 0.13	0.631 / 0.612	25	115	19 / 18	6.0 / 5.8	22	5	700	0.8	79	4.0	74.90	312	30.16	102	103	582.200	585.200	Stack 1	1.0000
076 BAI SO3 02	1335	1349	14-Jun-94	8.4	± 0.26	0.605 / 0.770		130	27 / 28	8.6 / 8.2	22	5	700	0.8	80	3.9	74.70	313	30.16	103	105	585.700	588.700	Stack 1	1.0000
076 BAI SO3 03	1403	1417	14-Jun-94	15.7	± 0.04	1.769 / 1.782		106	48 / 49	15.7 / 15.7	22	5	700	0.8	79	4.2	74.50	314	30.16	105	106	589.280	592.280	Stack 1	1.0000

Average SO3 Concentration, ppm(v), wet @3% O2: 12.1 ± 5.2
Testing Interval Averages: O2 4.0 σ 0.2; Load 79 σ 1

BAI SO2

Sample	Time Begin	Time End	Date 1994	ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm dry, @3% O2
076 BAI SO2 01	1307	1320	14-Jun-94	2007	± 38	4.384 / 4.268	200	703	6421 / 6251	2033.9 / 1980.1
076 BAI SO2 02	1335	1349	14-Jun-94	2854	± 32	4.661 / 4.780		837	8476 / 8335	2676.4 / 2631.8
076 BAI SO2 03	1403	1417	14-Jun-94	1651	± 4	3.692 / 3.705		685	5115 / 5133	1648.4 / 1654.2

Average SO2 Concentration, ppm(v), wet @3% O2: 2104 ± 508

BAO SO3

Sample	Time Begin	Time End	Date 1994	ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm dry, @3% O2	Condition Number	Flow scfm	Temp °F	NH3 NO	Unit 5 Boiler Load, MW	Sample O2 %	Volume Total STD l, dry	Temp Avg °K	Barometric Pressure In. Hg	Temp Begin °F	Temp End °F	Volume Begin ft³	Volume End ft³	Meter I.D	Corr. Factor
076 BAO SO3 01	1312	1325	14-Jun-94	8.6	± 0.03	0.828 / 0.824	25	94.8	20 / 20	8.7 / 8.6	22	5	700	0.8	79	8.2	74.46	311	30.16	100	100	46.600	49.600	Stack 2	0.9897
076 BAO SO3 02	1341	1355	14-Jun-94	9.8	± 0.12	1.220 / 1.200		74.3	24 / 23	9.9 / 9.8	22	5	700	0.8	80	8.1	74.46	311	30.16	100	100	50.100	53.100	Stack 2	0.9897
076 BAO SO3 03	1408	1423	14-Jun-94	11.4	± 0.06	1.037 / 1.045		101	27 / 27	11.4 / 11.5	22	5	700	0.8	80	8.0	74.39	311	30.16	100	101	53.400	56.400	Stack 2	0.9897

Average SO3 Concentration, ppm(v), wet @3% O2: 10.6 ± 1.1
Testing Interval Averages: O2 8.1 σ 0.1; Load 80 σ 1

BAO SO2

Sample	Time Begin	Time End	Date 1994	ppm(v), dry, @3% O2 Average	σ	SO4⁻ mg/ml	Dil	Vol. ml	SO4⁻ µM	SO3 or SO3 ppm dry, @3% O2
076 BAO SO2 01	1312	1325	14-Jun-94	2171	± 2	4.011 / 4.006	200	613	5122 / 5116	2172.0 / 2169.3
076 BAO SO2 02	1341	1355	14-Jun-94	2209	± 4	3.938 / 3.927		841	5258 / 5244	2211.3 / 2208.3
076 BAO SO2 03	1408	1423	14-Jun-94	2190	± 11	3.887 / 3.915		645	5223 / 5261	2182.4 / 2186.1

Average SO2 Concentration, ppm(v), wet @3% O2: 2190 ± 19

Table A-4, continued (Reference Table 3-2)

Number	Time, Begin	Time, End	Date, 1994	Conc. Avg ppm(v), dry, @3% O2	σ	SO4* mg/ml	Dil ml	Vol. ml	SO4* µM	SO2 or SO3 ppm dry @3% O2	Condition Number	Flow scfm	Temp °F	NH3/NO2	Unit 5 Boiler Load, MW	Sample O2 %	Volume Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
076 CAI SO3 01	1338	1350	13-Jun-94	15.3 ±	0.00	2.181	25	90.3	51	15.3	22	5	700	0.8	80	3.4	78.94	304	30.13	87	87	570.500	573.500	Stack 1	1.0000
						2.182			51	15.3															
076 CAI SO3 02	1403	1416	13-Jun-94	21.2 ±	0.07	2.628		104	71	21.1	22	5	700	0.8	80	3.3	78.80	304	30.13	87	89	574.000	577.000	Stack 1	1.0000
						2.640			72	21.2															
076 CAI SO3 03	1427	1442	13-Jun-94	20.1 ±	0.13	2.452		105	67	20.1	22	5	700	0.8	80	3.4	78.25	306	30.13	91	93	577.400	580.400	Stack 1	1.0000
						2.429			68	20.0															
			Testing Interval Averages												80	3.4									
			σ												0	0.1									
			Average SO3 Concentration, ppm(v), wet @ 3% O2	20.6 ±	0.8																				
076 CAI SO2 01	1338	1350	13-Jun-94	1860 ±	17	5.457	200	553	6287	1872.1															
						5.388			6205	1847.7															
076 CAI SO2 02	1403	1416	13-Jun-94	1805 ±	0	4.901		596	6085	1805.0															
						4.902			6087	1805.4															
076 CAI SO2 03	1427	1442	13-Jun-94	1802 ±	3	4.513		637	5989	1799.7															
						4.523			6002	1803.7															
			Average SO2 Concentration, ppm(v), wet @ 3% O2	1822 ±	33																				
076 CAO SO3 01	1342	1356	13-Jun-94	11.9 ±	0.05	1.535	25	88	34	11.8	22	5	700	0.8	80	5.4	75.28	307	30.13	94	93	35.900	38.900	Stack 2	0.8897
						1.544			35	11.0															
076 CAO SO3 02	1408	1421	13-Jun-94	13.2 ±	0.08	1.938		75	38	13.1	22	5	700	0.8	80	5.8	75.48	306	30.13	92	92	39.300	42.300	Stack 2	0.8897
						1.951			38	13.2															
076 CAO SO3 03	1435	1448	13-Jun-94	12.8 ±	0.07	1.376		103	37	12.8	22	5	700	0.8	80	5.6	75.46	306	30.13	92	92	42.600	45.600	Stack 2	0.8897
						1.365			37	12.7															
			Testing Interval Averages												80	5.5									
			σ												0	0.1									
			Average SO2 Concentration, ppm(v), wet @ 3% O2	13.0 ±	0.3																				
076 CAO SO2 01	1342	1356	13-Jun-94	1963.4 ±	3.7	4.474	200	612	5704	1960.8															
						4.466			5720	1966.0															
076 CAO SO2 02	1408	1421	13-Jun-94	1953.4 ±	9.1	4.286		632	5843	1959.8															
						4.258			5608	1947.0															
076 CAO SO2 03	1435	1448	13-Jun-94	1918.0 ±	9.3	3.841		875	5542	1924.6															
						3.814			5504	1911.4															
			Average SO2 Concentration, ppm(v), wet @ 3% O2	1945 ±	24																				

Table A-5 (Reference Table 3-3)

		Time, Begin	Time, End	Date, 1994	Average ppm(v), dry @3% O₁	±	σ	SO₄⁼ mg/ml	Dil ml	Vol. ml	SO₄⁼ µM	SO₂ or SO₃ ppm, dry @3% O2	Condition Number	Flow scfm	Temp, °F	NH₃, NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In.Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
084	AAI SO3 01	0958	1010	17-Oct-94	23.2	±	0.01	6.636 / 6.632	12.5	88.5	75 / 75	23.2 / 23.2	22	5000	700	0.8	80	4.3	77.77	300	30.18	80	82	673,500	676,500	Stack 1	0.9980
084	AAI SO3 02	1028	1040	17-Oct-94	22.9	±	1.97	4.108 / 4.637		131	70 / 79	21.5 / 24.3	22	5000	700	0.8	80	4.4	79.11	301	30.18	82	83	676,900	679,900	Stack 1	0.9980
084	AAI SO3 03	1055	1108	17-Oct-94	19.1	±	0.02	4.065 / 4.071		114	60 / 60	19.1 / 19.1	22	5000	700	0.8	80	4.5	77.34	302	30.18	83	85	680,200	683,200	Stack 1	0.9980
	Average SO₃ Concentration, ppm(v), wet @ 3% O2				21.0	±	2.7					Testing Interval Averages / σ					80 / 0	4.4 / 0.1									
084	AAI SO2 01	0958	1010	17-Oct-94	1954	±	0.7	3.812 / 3.814	200	792	6290 / 6293	1953.4 / 1954.5															
084	AAI SO2 02	1028	1040	17-Oct-94	1896	±	2.2	3.679 / 3.673		806	6178 / 6168	1897.8 / 1894.8															
084	AAI SO2 03	1055	1108	17-Oct-94	1821	±	8.6	3.432 / 3.455		603	5741 / 5780	1814.9 / 1827.1															
	Average SO₂ Concentration, ppm(v), wet @ 3% O2				1890	±	67																				
084	AAO SO3 01	0949	1001	17-Oct-94	18.8	±	0.07	3.894 / 3.874	12.5	99.0	50 / 50	18.8 / 18.7	22	5000	700	0.8	80	7.3	78.53	295	30.18	70	72	572,900	575,900	SRI 3	0.9891
084	AAO SO3 02	1020	1037	17-Oct-94	12.5	±	0.02	2.519 / 2.512		104	34 / 34	12.5 / 12.5	22	5000	700	0.8	80	7.0	78.46	295	30.18	71	72	576,500	579,500	SRI 3	0.9891
084	AAO SO3 03	1050	1105	17-Oct-94	15.4	±	0.06	3.079 / 3.660		65.3	41 / 41	15.4 / 15.3	22	5000	700	0.8	80	7.3	78.24	296	30.18	73	73	579,900	582,900	SRI 3	0.9891
	Average SO₃ Concentration, ppm(v), wet @ 3% O2				15.6	±	3.1					Testing Interval Averages / σ					80 / 0	7 / 0.2									
084	AAO SO2 01	0949	1001	17-Oct-94	1358	±	5.9	2.783 / 2.766	200	628	3629 / 3607	1362.8 / 1354.2															
084	AAO SO2 02	1020	1037	17-Oct-94	1429	±	4.8	2.935 / 2.921		637	3895 / 3876	1432.0 / 1425.2															
084	AAO SO2 03	1050	1105	17-Oct-94	1451	±	1.8	2.819 / 2.814		656	3853 / 3846	1451.8 / 1449.2															
	Average SO₂ Concentration, ppm(v), wet @ 3% O2				1413	±	48																				

A-13

Table A-5, continued (Reference Table 3-3)

Concentration / Analytical

		Time, Begin	Time, End	Date, 1994	ppm(v), dry, @3% O2 Average	σ	SO4° mg/ml	Dil	Vol. ml	SO4° µM	SO2 or SO3 ppm, dry, @3% O2
093 BAI SO3 01		1307	1322	11-Oct-94	4.8	± 0.01	1.285 / 1.283	12.5	98.3	16 / 16	4.8 / 4.8
093 BAI SO3 02		1347	1400	11-Oct-94	4.6	± 0.01	1.169 / 1.167		103	16 / 16	4.6 / 4.6
093 BAI SO3 03		1415	1429	11-Oct-94	3.0	± 0.05	0.738 / 0.722		108	10 / 10	3.0 / 3.0
Average SO3 Concentration, ppm(v), wet @ 3% O2					3.8	± 1.2					

Reactor Conditions / Dry Gas Meter (BAI SO3)

Condition Number	Flow scfm	Temp °F	NH3/NOx	Unit 5 Boiler Load, MW	Sample O2 %	Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
22	5000	700	0.8	88	3.4	78.47	296	30.00	73	73	579.200	582.200	Stack 1	0.998
22	5000	700	0.8	88	3.6	76.47	296	30.00	73	73	582.600	585.600	Stack 1	0.998
22	5000	700	0.8	89	3.7	78.40	296	30.00	73	74	585.800	588.800	Stack 1	0.998
Testing Interval Averages				88	3.6									
σ				1	0.2									

Analytical (BAI SO2)

		Time, Begin	Time, End	Date, 1994	ppm(v), dry, @3% O2 Average	σ	SO4° mg/ml	Dil	Vol. ml	SO4° µM	SO2 or SO3 ppm, dry, @3% O2
093 BAI SO2 01		1307	1322	11-Oct-94	2002	± 0.0	4.284 / 4.284	200	772	6858 / 6858	2002.4 / 2002.4
093 BAI SO2 02		1347	1400	11-Oct-94	2028	± 5.8	4.201 / 4.218		783	6853 / 6881	2024.0 / 2032.2
093 BAI SO2 03		1415	1429	11-Oct-94	2064	± 7.6	4.425 / 4.402		755	6960 / 6924	2069.6 / 2058.9
Average SO2 Concentration, ppm(v), wet @ 3% O2					2032	± 31					

Concentration / Analytical (BAO SO3)

		Time, Begin	Time, End	Date, 1994	ppm(v), dry, @3% O2 Average	σ	SO4° mg/ml	Dil	Vol. ml	SO4° µM	SO2 or SO3 ppm, dry, @3% O2
093 BAO SO3 01		1300	1315	11-Oct-94	11.7	± 0.02	2.905 / 2.913	13	73.9	28 / 28	11.7 / 11.7
093 BAO SO3 02		1340	1354	11-Oct-94	8.4	± 0.03	1.841 / 1.848		100	21 / 21	8.3 / 8.4
093 BAO SO3 03		1408	1425	11-Oct-94	8.8	± 0.02	1.497 / 1.492		118	23 / 23	8.8 / 8.8
Average SO3 Concentration, ppm(v), wet @ 3% O2					9.6	± 1.8					

Reactor Conditions / Dry Gas Meter (BAO SO3)

Condition Number	Flow scfm	Temp °F	NH3/NOx	Unit 5 Boiler Load, MW	Sample O2 %	Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
22	5000	700	0.8	88	8.7	78.73	292	30.00	65	68	472.300	475.300	SRI 3	0.9891
22	5000	700	0.8	88	7.8	78.38	294	30.00	67	71	476.000	479.000	SRI 3	0.9891
22	5000	700	0.8	89	7.8	78.51	293	30.00	68	68	479.600	482.600	SRI 3	0.9891
Testing Interval Averages				88	8									
σ				1	1									

Analytical (BAO SO2)

		Time, Begin	Time, End	Date, 1994	ppm(v), dry, @3% O2 Average	σ	SO4° mg/ml	Dil	Vol. ml	SO4° µM	SO2 or SO3 ppm, dry, @3% O2
093 BAO SO2 01		1300	1315	11-Oct-94	1729	± 2.1	3.515 / 3.509	200	568	4145 / 4138	1730.2 / 1727.2
093 BAO SO2 02		1340	1354	11-Oct-94	1817	± 4.5	2.858 / 2.848		783	4859 / 4843	1819.8 / 1813.4
093 BAO SO2 01		1408	1425	11-Oct-94	1633	± 6.8	2.883 / 2.866		710	4284 / 4239	1637.6 / 1627.9
Average SO2 Concentration, ppm(v), wet @ 3% O2					1726	± 92					

Table A-5, continued (Reference Table 3-3)

CAI SO3

Stream	Time Begin	Time End	Date, 1994	ppm(v), dry @ 3% O2 Average	σ	SO4⁻ (mg/ml)	Dil (ml)	Vol (ml)	SO4⁻ (µM)	SO2 or SO3 (ppm, dry @ 3% O2)	Condition Number	Flow (scfm)	Temp (°F)	NH3/NO2	Unit 5 Boiler Load (MW)	Sample O2 (%)	Volume Total STD l, dry	Temp Avg (°K)	Barometric Pressure (in. Hg)	Temp Begin (°F)	Temp End (°F)	Volume Begin (ft³)	Volume End (ft³)	Meter I.D	Corr. Factor
093 CAI SO3 01	0931	0947	10-Oct-94	3.4	0.01	0.896 / 0.888	12.5	101	11.8 / 11.8	3.4 / 3.4	22	5000	700	0.8	88	2.8	77.87	297	29.94	77	75	548.200	549.200	Stack 1	0.998
093 CAI SO3 02	1003	1017	10-Oct-94	4.7	0.06	1.076 / 1.094	118		16.5 / 16.8	4.7 / 4.7	22	5000	700	0.8	88	2.7	78.17	298	29.94	74	74	549.700	552.700	Stack 1	0.998
093 CAI SO3 03	1041	1056	10-Oct-94	3.5	0.04	0.860 / 0.874	111		12.4 / 12.6	3.5 / 3.6	22	5000	700	0.8	89	2.7	78.31	298	29.94	73	73	553.000	556.000	Stack 1	0.998

Average SO3 Concentration, ppm(v), wet @ 3% O2: 4.1 ± 0.8
Testing Interval Averages: 88, 2.7 ; σ 1, 0.1

CAI SO2

Stream	Time Begin	Time End	Date, 1994	Average	σ	SO4⁻ (mg/ml)	Dil (ml)	Vol (ml)	SO4⁻ (µM)	SO2 or SO3 (ppm, dry @ 3% O2)
093 CAI SO2 01	0931	0947	10-Oct-94	1964	2.7	4.176 / 4.184	200	783	6899 / 6912	1962.5 / 1966.3
093 CAI SO2 02	1003	1017	10-Oct-94	1938	6.1	4.312 / 4.293	787		6890 / 6860	1942.0 / 1933.4
093 CAI SO2 03	1041	1056	10-Oct-94	1944	0.3	4.441 / 4.440	747		6911 / 6910	1944.2 / 1943.8

Average SO2 Concentration, ppm(v), wet @ 3% O2: 1949 ± 14

CAO SO3

Stream	Time Begin	Time End	Date, 1994	Average	σ	SO4⁻ (mg/ml)	Dil (ml)	Vol (ml)	SO4⁻ (µM)	SO2 or SO3 (ppm, dry @ 3% O2)	Condition Number	Flow (scfm)	Temp (°F)	NH3/NO2	Unit 5 Boiler Load (MW)	Sample O2 (%)	Volume Total STD l, dry	Temp Avg (°K)	Barometric Pressure (in. Hg)	Temp Begin (°F)	Temp End (°F)	Volume Begin (ft³)	Volume End (ft³)	Meter I.D	Corr. Factor
093 CAO SO3 01	0926	0941	10-Oct-94	7.6	0.05	1.888 / 2.005	12.5	97.8	25.3 / 25.5	7.5 / 7.6	22	5000	700	0.8	88	3.5	77.54	296	29.94	73	74	437.400	440.400	SRI 3	0.888
093 CAO SO3 02	0957	1013	10-Oct-94	7.6	0.01	1.707 / 1.703	116		25.6 / 25.7	7.6 / 7.6	22	5000	700	0.8	89	3.5	77.98	294	29.94	71	70	441.100	444.100	SRI 3	0.888
093 CAO SO3 03	1038	1050	10-Oct-94	6.6	0.01	1.716 / 1.720	101		22.6 / 22.6	6.6 / 6.7	22	5000	700	0.8	89	3.5	78.35	293	29.94	88	88	445.200	448.200	SRI 3	0.888

Average SO3 Concentration, ppm(v), wet @ 3% O2: 7.3 ± 0.5
Testing Interval Averages: 89, 4 ; σ 1, 0

CAO SO2

Stream	Time Begin	Time End	Date, 1994	Average	σ	SO4⁻ (mg/ml)	Dil (ml)	Vol (ml)	SO4⁻ (µM)	SO2 or SO3 (ppm, dry @ 3% O2)
093 CAO SO2 01	0926	0941	10-Oct-94	1515	0.8	3.996 / 3.999	200	612	5095 / 5099	1514.1 / 1515.2
093 CAO SO2 02	0957	1013	10-Oct-94	1427	3.5	3.703 / 3.690	627		4837 / 4820	1429.4 / 1424.3
093 CAO SO2 03	1038	1050	10-Oct-94	1495	8.0	3.346 / 3.365	727		5088 / 5097	1490.5 / 1498.9

Average SO2 Concentration, ppm(v), wet @ 3% O2: 1479 ± 46

Table A-6 (Reference Table 3-4)

	Time, Begin	Time, End	Date	Concentration ppm(v), dry, @3% O₂ Average		σ	Analytical Cl⁻ mg/ml	Dil	Vol. ml	Cl⁻ µM	HCl ppm dry @3% O₂	Reactor Conditions Condition Number	Flow scfm	Temp, °F	NH₃/NO₂	Unit 5 Boiler Load, MW	Sample O₂ %	Dry Gas Meter Volume, Total STD l, dry	Temp, Avg. °K	Barometric Pressure In. Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
094 AAI HCl 01	1309	1323	17-Oct	92.4	±	0.32	1.428 / 1.421	25	783	291 / 290	92.6 / 92.2	22	5000	700	0.8	80	4.5	76.85	304	30.18	88	87	683.500	686.500	Stack 1	0.998
094 AAI HCl 02	1334	1347	17-Oct	94.2	±	0.45	1.335 / 1.344		853	297 / 299	93.9 / 94.5	22	5000	700	0.8	80	4.4	76.78	304	30.18	88	88	686.900	689.900	Stack 1	0.998
094 AAI HCl 03	1356	1408	17-Oct	91.5	±	0.40	1.610 / 1.600		680	285 / 283	91.8 / 91.2	22	5000	700	0.8	80	4.6	76.43	306	30.18	90	91	690.200	693.200	Stack 1	0.998
Average HCl Concentration, ppm(v), wet @3% O2				93	±	1.9						Testing Interval Averages	5000	700	0.8	80	4.5									
												σ				0	0.1									
093 AAO HCl 01	1304	1319	17-Oct	82.1	±	6.40	0.986 / 1.101	25	790	203 / 227	77.6 / 86.6	22	5000	700	0.8	80	7.4	77.65	298	30.18	76	78	583.300	586.300	SRI 3	0.9891
093 AAO HCl 02	1328	1343	17-Oct	80.9	±	0.69	1.160 / 1.146		697	211 / 208	81.4 / 80.4	22	5000	700	0.8	80	7.5	77.44	299	30.18	78	79	586.900	589.900	SRI 3	0.9891
093 AAO HCl 03	1350	1404	17-Oct	80.2	±	0.52	1.098 / 1.088		735	210 / 208	80.6 / 79.9	22	5000	700	0.8	80	7.4	77.44	299	30.18	78	79	590.300	593.300	SRI 3	0.9891
Average HCl Concentration, ppm(v), wet @3% O2				81	±	1.0						Testing Interval Averages	5000	700	0.8	80	7									
												σ				0	0									

Table A-6, continued (Reference Table 3-4)

Sample ID	Time, Begin	Time, End	Date	Conc. Average	±	σ	Cl⁻ mg/ml	Dil	Vol. ml	Cl⁻ μM	HCl ppm, dry @3% O₂	Condition Number	Flow scfm	Temp. °F	NH₃ NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Vol. Total STD l, dry	Temp. Avg. °K	Barometric Pressure In.Hg	Temp. Begin °F	Temp. End °F	Volume Begin ft³	Volume End ft³	Meter I.D	Corr. Factor
093 BAI HCl 01	0954	1006	11-Oct	105.2	±	6.98	1.786	25	805	374	110.2	22	5000	700	0.8	88	3.5	78.32	296	30.00	74	74	568.900	571.900	Stack 1	0.998
							1.626			341	100.3															
093 BAI HCl 02	1018	1031	11-Oct	97.6	±	0.15	1.421		908	336	97.7	22	5000	700	0.8	89	3.3	78.32	296	30.00	74	74	572.300	575.300	Stack 1	0.998
							1.418			335	97.5															
093 BAI HCl 03	1046	1059	11-Oct	104.8	±	0.61	1.695		810	358	105.2	22	5000	700	0.8	89	3.5	78.32	296	30.00	74	74	575.800	578.800	Stack 1	0.998
							1.681			355	104.3															
Average HCl Concentration, ppm(v), wet @ 3% O₂				101	±	5.0						Testing Interval Averages				89	3.4									
															σ	1	0.1									
093 BAO HCl 01	0947	1002	11-Oct	88.3	±	0.46	1.205	25	740	232	88.0	22	5000	700	0.8	88	7.4	78.36	294	30.00	68	70	461.200	464.200	SRI 3	0.9891
							1.214			234	88.7															
093 BAO HCl 02	1012	1028	11-Oct	95.0	±	6.92	1.317		774	265	99.9	22	5000	700	0.8	89	7.3	78.36	294	30.00	68	70	464.800	467.800	SRI 3	0.9891
							1.188			239	90.1															
093 BAO HCl 03	1041	1055	11-Oct	96.0	±	1.77	1.095		753	215	94.7	22	5000	700	0.8	88	9.3	78.36	294	30.00	70	70	468.500	471.500	SRI 3	0.9891
							1.124			220	97.2															
Average HCl Concentration, ppm(v), wet @ 3% O₂				93	±	4.1						Testing Interval Averages				88	8									
															σ	1	1									

Concentration: ppm(v), dry, @3% O₂ • Cl⁻ μM • HCl ppm, dry @3% O₂

Table A-6, continued (Reference Table 3-4)

	Time, Begin	Time, End	Date	Concentration ppm(v), dry, @3% O₂ Average	σ	Cl⁻ mg/ml	Dil ml	Vol. ml	Cl⁻ µM	HCl ppm, dry, @3% O₂	Condition Number	Flow scfm	Temp, °F	NH₃/NOₓ	Unit 5 Boiler Load, MW	Sample O₂ %	Volume, Total STD1, dry	Temp, Avg. °K	Barometric Pressure in.Hg	Temp, Begin °F	Temp, End °F	Volume, Begin ft³	Volume, End ft³	Meter I.D	Corr. Factor
093 CAI HCI 01	1329	1343	10-Oct	106.5	± 11.72	2.009 / 1.719	25	777	407 / 348	114.8 / 98.2	22	5000	700	0.8	89	2.6	77.58	299	29.94	79	77	556.200	559.200	Stack 1	0.998
093 CAI HCI 02	1359	1412	10-Oct	96.9	± 2.42	1.612 / 1.670		807	339 / 351	95.2 / 98.7	22	5000	700	0.8	89	2.6	77.95	297	29.94	76	75	559.700	562.700	Stack 1	0.998
093 CAI HCI 03	1424	1436	10-Oct	102.6	± 0.31	1.644 / 1.637		846	362 / 361	102.8 / 102.4	22	5000	700	0.8	88	2.8	78.02	297	29.94	75	75	563.100	566.100	Stack 1	0.998
Testing Interval Averages															89	2.7									
σ															1	0.1									
Average HCI Concentration, ppm(v), wet @ 3% O2				100	± 4.0																				
093 CAO HCI 01	1324	1339	10-Oct	85.2	± 0.95	1.540 / 1.516	25	720	289 / 284	85.9 / 84.6	22	5000	700	0.8	89	3.5	77.47	296	29.94	74	74	449.200	452.200	SRI 3	0.9891
093 CAO HCI 02	1353	1408	10-Oct	85.5	± 1.15	1.516 / 1.545		725	286 / 292	84.7 / 86.4	22	5000	700	0.8	89	3.4	77.40	297	29.94	74	75	453.500	456.500	SRI 3	0.9891
093 CAO HCI 03	1418	1434	10-Oct	83.9	± 0.73	1.554 / 1.535		701	284 / 280	84.5 / 83.4	22	5000	700	0.8	89	3.5	77.40	297	29.94	74	75	457.400	460.400	SRI 3	0.9891
Testing Interval Averages															89	3									
σ															0	0									
Average HCI Concentration, ppm(v), wet @ 3% O2				85	± 0.8																				

Table A-7 (Reference Table 3-5)

INPUT DATA			
DATE:	6/9/94		
RUN ID:	**075-AAO-01**		
Flue gas O2	8.3 % wet	Nozzle diameter	0.219 in.
Flue gas CO2	10.7 % wet	Gas meter corr.	1.002
Flue gas N2	72.32 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	8.68 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in. Hg		
Stack Pressure	-22.5 in. H2O		
Stack Temp	338.67 °F		
Volume water	104 ml	Meter volume	54.666 ft3
Particle mass	8542.8 mg	SQRT DP pitot	0.9326
Run time	90 min	AVG DH orifice	1.31 in. H2O
Meter Temp	104.61 °F		

FINAL CALCULATED DATA			
% WATER	8.68%	gr/ACF :	1.4629 wet
% ISOKINETIC	98.69%	gr/SCF :	2.5970 dry
STACK VEL.	64.43 ft/s	mg/ACM :	3347.80 wet
		mg/SCM :	5942.89 dry
STACK FLOW	12989 ACFM (wet)		
STACK FLOW	7295 SCFM (dry)	LB/MBTU	5.9336

INPUT DATA

DATE: 6/9/94
RUN ID: 075-AAO-02

Flue gas O2	8.3 % wet	Nozzle diameter	0.219 in.
Flue gas CO2	10.7 % wet	Gas meter corr.	1.002
Flue gas N2	72.24 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	8.76 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in. Hg		
Stack Pressure	-22.5 in. H2O		
Stack Temp	339 °F		
Volume water	104.2 ml	Meter volume	54.551 ft3
Particle mass	7693.8 mg	SQRT DP pitot	0.9518
Run time	90 min	AVG DH orifice	1.302 in. H2O
Meter Temp	108.°F		

FINAL CALCULATED DATA

% WATER	8.76%	gr/ACF :	1.3266 wet
% ISOKINETIC	96.00%	gr/SCF :	2.3584 dry
STACK VEL.	65.78 ft/s	mg/ACM :	3035.86 wet
		mg/SCM :	5396.93 dry
STACK FLOW	13262 ACFM (wet)		
STACK FLOW	7438 SCFM (dry)	LB/MBTU	5.3885

INPUT DATA

DATE: 6/9/94
RUN ID: 075-AAO-03

Flue gas O2	8.3 % wet	Nozzle diameter	0.219 in.
Flue gas CO2	10.7 % wet	Gas meter corr.	1.002
Flue gas N2	72.2 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	8.8 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in. Hg		
Stack Pressure	-22.5 in. H2O		
Stack Temp	337.89 °F		
Volume water	106.2 ml	Meter volume	55.522 ft3
Particle mass	8237.2 mg	SQRT DP pitot	0.9303
Run time	90 min	AVG DH orifice	1.304 in. H2O
Meter Temp	110.278 °F		

FINAL CALCULATED DATA

% WATER	8.80%	gr/ACF :	1.4023 wet
% ISOKINETIC	99.55%	gr/SCF :	2.4903 dry
STACK VEL.	64.25 ft/s	mg/ACM :	3209.06 wet
		mg/SCM :	5698.71 dry
STACK FLOW	12953 ACFM (wet)		
STACK FLOW	7272 SCFM (dry)	LB/MBTU	5.6898

INPUT DATA

DATE: 6/10/94
RUN ID: 075-BAO-01

Flue gas O2	7.8 % wet	Nozzle diameter	0.219 in.
Flue gas CO2	13 % wet	Gas meter corr.	1.002
Flue gas N2	71.24 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	7.96 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.04 in. Hg		
Stack Pressure	-21 in. H2O		
Stack Temp	317.56 °F		
Volume water	102 ml	Meter volume	58.302 ft3
Particle mass	8454.9 mg	SQRT DP pitot	0.9984
Run time	90 min	AVG DH orifice	1.5 in. H2O
Meter Temp	99.78 °F		

FINAL CALCULATED DATA

% WATER	7.96%	gr/ACF :	1.3982 wet
% ISOKINETIC	97.58%	gr/SCF :	2.3858 dry
STACK VEL.	67.44 ft/s	mg/ACM :	3199.69 wet
		mg/SCM :	5459.75 dry
STACK FLOW	13596 ACFM (wet)		
STACK FLOW	7936 SCFM (dry)	LB/MBTU	5.2431

INPUT DATA

DATE: 6/10/94
RUN ID: 075-BAO-02

Flue gas O_2	7.8 % wet	Nozzle diameter	0.219 in.
Flue gas CO_2	13 % wet	Gas meter corr.	1.002
Flue gas N_2	71.22 % wet	Pitot corr. (C_p)	0.821
Flue gas H_2O	7.98 % wet	Duct Area	3.36 ft^2
Ambient Pressure	30.04 in. Hg		
Stack Pressure	-21.0 in. H_2O		
Stack Temp	314 °F		
Volume water	101.3 ml	Meter volume	58.259 ft^3
Particle mass	8093.5 mg	SQRT DP pitot	1.0089
Run time	90 min	AVG DH orifice	1.53 in. H_2O
Meter Temp	105 °F		

FINAL CALCULATED DATA

% WATER	7.98%	gr/ACF :	1.3579 wet
% ISOKINETIC	95.41%	gr/SCF :	2.3062 dry
STACK VEL.	67.99 ft/s	mg/ACM :	3107.35 wet
		mg/SCM :	5277.62 dry
STACK FLOW	13706 ACFM (wet)		
STACK FLOW	8038 SCFM (dry)	LB/MBTU	5.0682

A-23

INPUT DATA

DATE: 6/10/94
RUN ID: 075-BAO-03

Flue gas O2	7.8 % wet	Nozzle diameter	0.219 in.
Flue gas CO2	13 % wet	Gas meter corr.	1.002
Flue gas N2	71.24 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	7.96 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.04 in. Hg		
Stack Pressure	-21 in. H2O		
Stack Temp	316.55 °F		
Volume water	104.6 ml	Meter volume	60.638 ft3
Particle mass	8464 mg	SQRT DP pitot	1.0552
Run time	90 min	AVG DH orifice	1.661 in. H2O
Meter Temp	107.83 °F		

FINAL CALCULATED DATA

% WATER	7.96%	gr/ACF :	1.3664 wet
% ISOKINETIC	94.61%	gr/SCF :	2.3285 dry
STACK VEL.	71.23 ft/s	mg/ACM :	3126.94 wet
		mg/SCM :	5328.59 dry
STACK FLOW	14360 ACFM (wet)		
STACK FLOW	8393 SCFM (dry)	LB/MBTU	5.1172

TESTING AND ANALYTICAL SERVICES FOR THE INNOVATIVE CLEAN COAL TECHNOLOGY DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION (SCR) TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO_x) EMISSIONS FROM HIGH SULFUR COAL

Final Report for Task 4

Air Heater Tests
Reactors A, B, C
First Parametric Series

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September 1, 1995

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

ACKNOWLEDGMENTS

Section 1

INTRODUCTION

This is the fifth in a series of reports describing the results of testing and analytical services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The test results for the initial set of air heater tests are summarized in this report. The air heater tests were conducted during the last week of April and the first two weeks of May, 1994, following the completion of the first sequence of parametric testing under Task 4: Long-Term Parametric Tests.

The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The test facility includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 °F down to 300 °F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C's air heater incorporates a heat pipe design.

The original design of the SCR test facility also included bypass heat exchangers on each of the three large reactors. These units were intended for use during parametric testing of the reactors (Task 4: Long Term Parametric Tests) so that flue gas containing higher concentrations of slip ammonia could be diverted around the air heaters. However, the bypass heat exchangers did not function as planned and the practice of bypassing the air heaters on Reactors A, B and C during parametric testing was abandoned. Thus, the air heater test data presented in this report include the effects of day-long periods of exposure to ammonia concentrations normally ranging from a few parts per million by volume ppm(v) to 20 ppm(v), although brief excursions approached 100 ppm(v).

The air heater testing included the determinations of particulate mass concentration and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia. All tests were conducted with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. In general, simultaneous tests were performed at the inlet and outlet of each of the three air heaters (except for mass concentration measurements at the outlet of the Reactor C air heater). The fly ash (particulate) catches from the mass concentration measurements were further analyzed in the laboratory to determine particle size distributions, ash mineralogy, and ash resistivity.

This report is divided into several sections. Section 2 describes the test methods used for the air heater testing. Section 3 reviews all of the test results. The test data are briefly summarized in Section 4. Tables containing original data summaries that were produced for the air heater testing are contained in Appendix A.

Section 2

TEST METHODS

A variety of test methods were used to characterize air heater performance. The following subsections briefly describe the test methods for ammonia (NH_3), sulfur dioxide (SO_2), sulfur trioxide (SO_3), hydrogen chloride (HCl), and particulate mass concentration, particle size, ash mineralogy, and ash resistivity. Figure 2-1 shows a sketch of a large reactor and its air heater (not to scale). The test port locations for air heater inlet and outlet measurements are indicated on the diagram. Test ports (three horizontal ports) upstream of catalyst layer 4 (the normal reactor outlet test location) were used to measure concentrations of air heater inlet particulate and gas constituents. Three test ports (horizontal) were installed in a transition piece in the outlet ducting of the Reactor A and B air heaters that were suitable for particulate and gas phase flue gas constituent testing. Test ports at this same location in the Reactor C air heater outlet were not installed. However, a single test port (horizontal) in a section of ducting at the inlet transition to the cyclone on Reactor C (downstream of the reactor bypass duct) was suitable for testing gas phase flue gas constituents. Turbulent flow at this location precluded mass concentration tests.

Air heater tests were performed with the reactors operating at Test Condition 22, the normal baseline or long-term operating condition. These operating conditions included a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 0.8, and a flue gas flow rate of 5000 wscfm. During measurements of air heater inlet and outlet ammonia concentrations only, additional tests were performed at Test Condition 24 (except Reactor A). The operating parameters for these tests were a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 1.0, and a flue gas flow rate of 5000 wscfm.

AMMONIA

Ammonia concentrations were measured simultaneously at the air heater inlet (upstream of catalyst layer 4) and at the air heater outlet on each of the large reactors. At each site and at each test condition, three independent tests were performed to determine an average ammonia concentration. Sampling ports located upstream of catalyst layer 4 were used to measure air heater inlet ammonia concentration. Flue gas was sampled at three equidistant points in each of the three test ports (a total of nine traverse points) to determine an average inlet ammonia concentration for each test. At the air heater outlet sampling location on Reactors A and B the probe also sampled at three equidistant points in each of the three test ports (a total of nine traverse points). At the single Reactor C outlet test port, the ammonia probe traversed three, equally-spaced, points (front to rear) during each test.
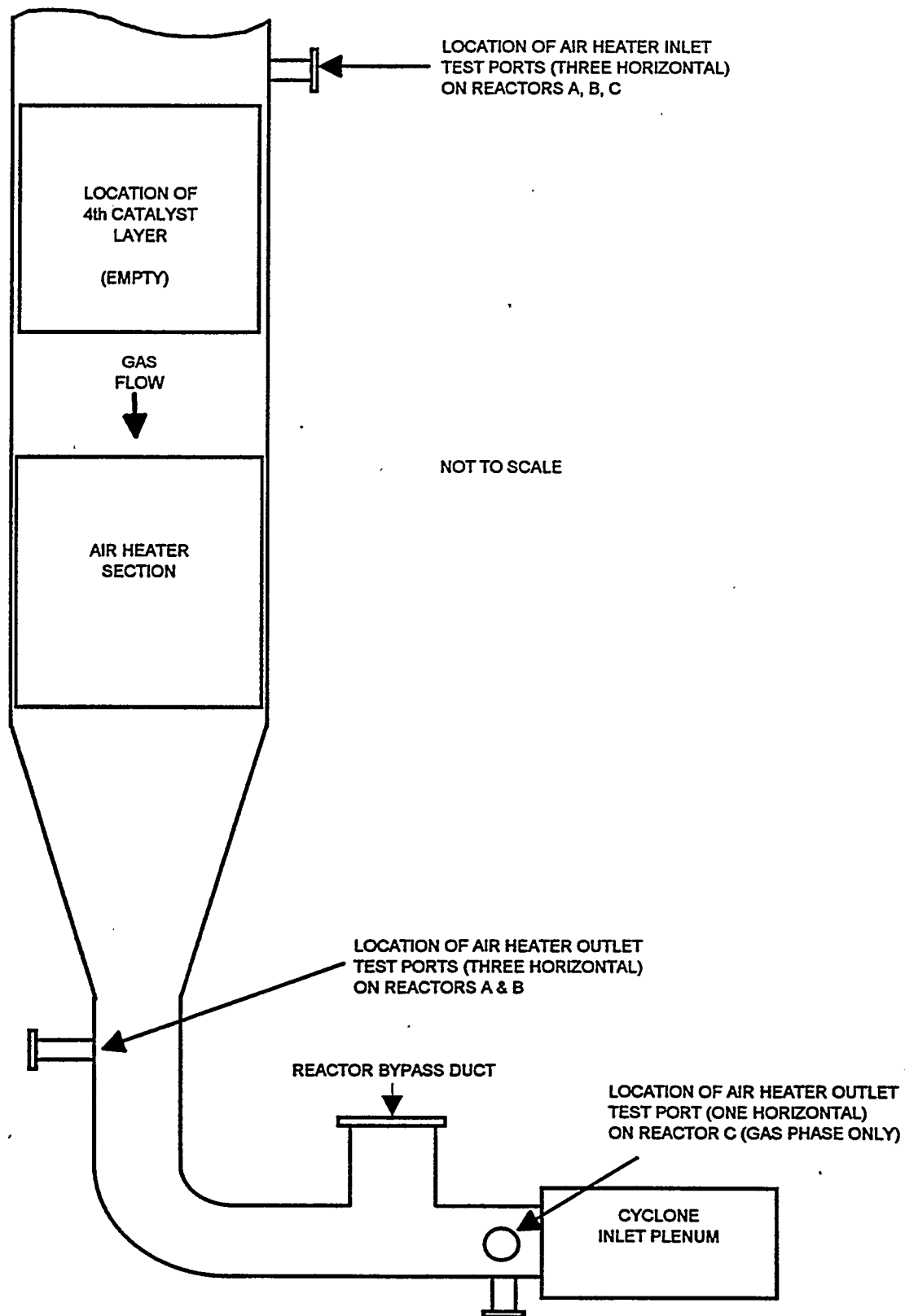
LOCATION OF AIR HEATER INLET
TEST PORTS (THREE HORIZONTAL)
ON REACTORS A, B, C

LOCATION OF
4th CATALYST
LAYER

(EMPTY)

GAS
FLOW

NOT TO SCALE

AIR HEATER
SECTION

LOCATION OF AIR HEATER OUTLET
TEST PORTS (THREE HORIZONTAL)
ON REACTORS A & B

REACTOR BYPASS DUCT

LOCATION OF AIR HEATER OUTLET
TEST PORT (ONE HORIZONTAL)
ON REACTOR C (GAS PHASE ONLY)

CYCLONE
INLET PLENUM

Figure 2-1. Schematic drawing of SCR reactor and air heater cross section (side view).

2-2

Figure 2-2. Schematic diagram of the ammonia sampling train.

A schematic drawing of the ammonia sampling train is shown in Figure 2-2. The flue gas sample is drawn through a glass-lined probe and a heated filter to remove particulate. The filter is maintained at the nominal flue gas temperature. The gas sample passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 0.1 normal solution of sulfuric acid. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. A second glass wool plug is placed in the line between the last two impingers to collect any ammonia aerosols that may escape the second impinger. The remainder of the train consists of a silica gel column to remove the last traces of water from the flue gas sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The samples were segregated during sample collection into a solid-phase sample and a gas-phase sample. The solid-phase sample consisted of the heated filter, all of the collected particulate, and the probe-wash liquid. The gas-phase sample consisted of the impinger liquids and the wash and rinse liquids. The two samples were analyzed separately to characterize the ammonia partitioning between the gas and solid phases.

At the outset of the Task 4 parametric testing, modified Greenberg-Smith type impingers were used for ammonia sampling. However, 100 mL gas sampling impingers, each containing about 50 mL of solution, were used for the air heater ammonia testing to reduce the detection limit to less than 1 ppm(v) for a reasonable sample volume, about 3 ft^3 of gas. The lower detection limit was needed for much of the air heater ammonia testing because ammonia concentrations are very low at the reactor exit when operating at the reactor design condition with an NH_3/NO_x ratio of 0.8.

The impinger solutions are made alkaline in the laboratory (converting the NH_4^+ ion to free NH_3 in solution). The concentration of ammonia is then determined with an ammonia ion specific electrode, Orion Model 920A. As mentioned above, three individual measurements of ammonia concentration are made for each test condition. Two independent determinations of the ammonia concentration are then conducted on each sample.

SULFUR DIOXIDE AND SULFUR TRIOXIDE

During the air heater evaluation, SO_2 and SO_3 concentrations were measured simultaneously at the reactor outlet upstream of catalyst layer 4 (air heater inlet) and the air heater outlet. Sulfur dioxide (SO_2) and sulfur trioxide (SO_3) were collected in a controlled condensation sampling train. All

2-3

tests were conducted using single-point sampling in either the right-hand port (air heater inlet test location) or middle port (air heater outlet test location).
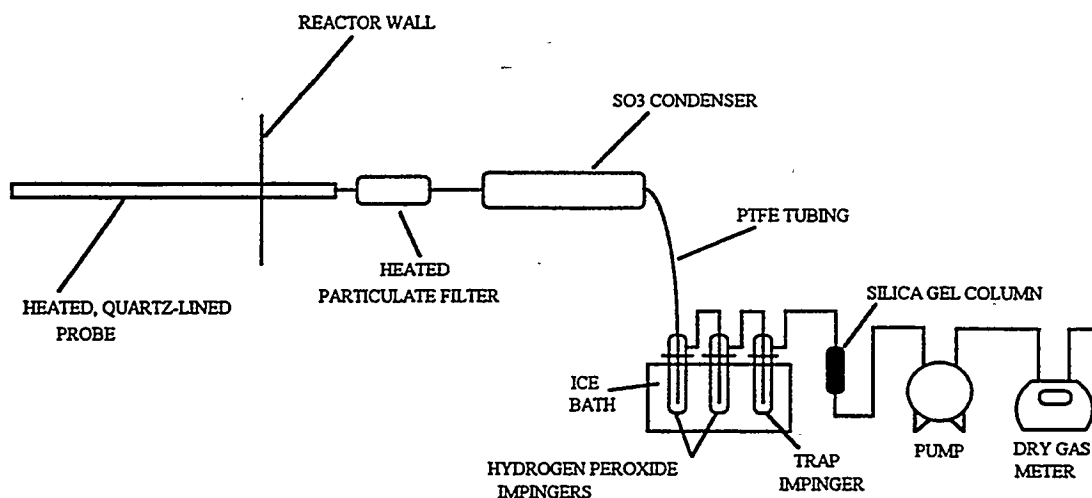
REACTOR WALL

SO3 CONDENSER

PTFE TUBING

HEATED PARTICULATE FILTER

SILICA GEL COLUMN

HEATED, QUARTZ-LINED PROBE

ICE BATH

HYDROGEN PEROXIDE IMPINGERS

TRAP IMPINGER

PUMP

DRY GAS METER

Figure 2-3. Schematic diagram of the controlled condensation sampling train for SO_2 and SO_3.

A schematic drawing of the SO_2/SO_3 sampling train is shown in Figure 2-3. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The next element in the train is the SO_3 condenser. The condenser is a length of quartz tubing packed with quartz wool and maintained between 120 °F and 130 °F in a heated water bath. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 3% solution of hydrogen peroxide to oxidize SO_2. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The SO_3 is collected in the condensing element and the SO_2 is collected in the bubblers by oxidation with the hydrogen peroxide, converting it to H_2SO_4 in solution. In the condenser the SO_3 present begins a hydration reaction with the water vapor present making H_2SO_4. The excess water vapor also condenses to produce a condensate of concentrated aqueous H_2SO_4. Thus, two solutions of H_2SO_4 are collected; one a very concentrated solution of limited amount containing the original SO_3 and the other a relatively weak solution in far greater amount containing the original SO_2. The concentrations of the sulfate ion are determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph.
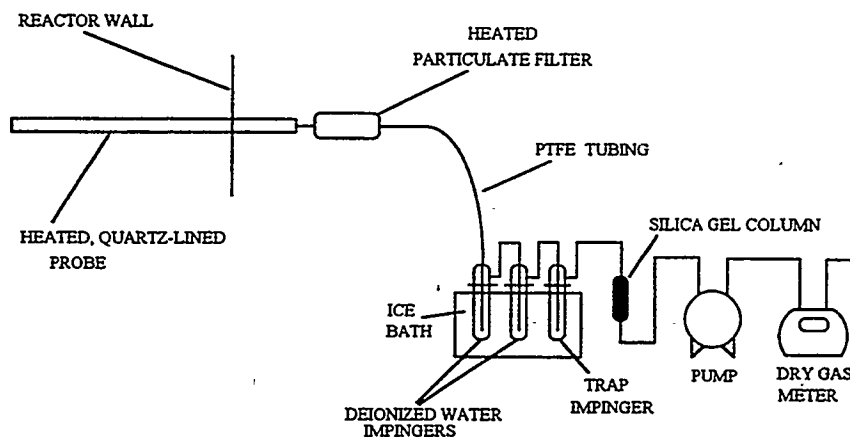
HYDROGEN CHLORIDE

Figure 2-4. Schematic diagram of the chloride sampling train.

A schematic sketch of the chloride sampling train is shown in Figure 2-4. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain type 1 de-ionized water for removal of HCl vapor. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The chloride ion concentration is determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph. Hydrogen chloride concentrations were determined by single-point sampling (at the mid-point of the duct) at the air heater inlet (upstream of the 4th catalyst layer, right-hand port) and at the air heater outlet (middle port). Inlet and outlet tests were conducted simultaneously.

MASS CONCENTRATION

Mass concentration was measured using a sampling train similar to that defined in the EPA Reference Test Method 17. A schematic drawing of the train used for mass concentration sampling is shown below in Figure 2-5.
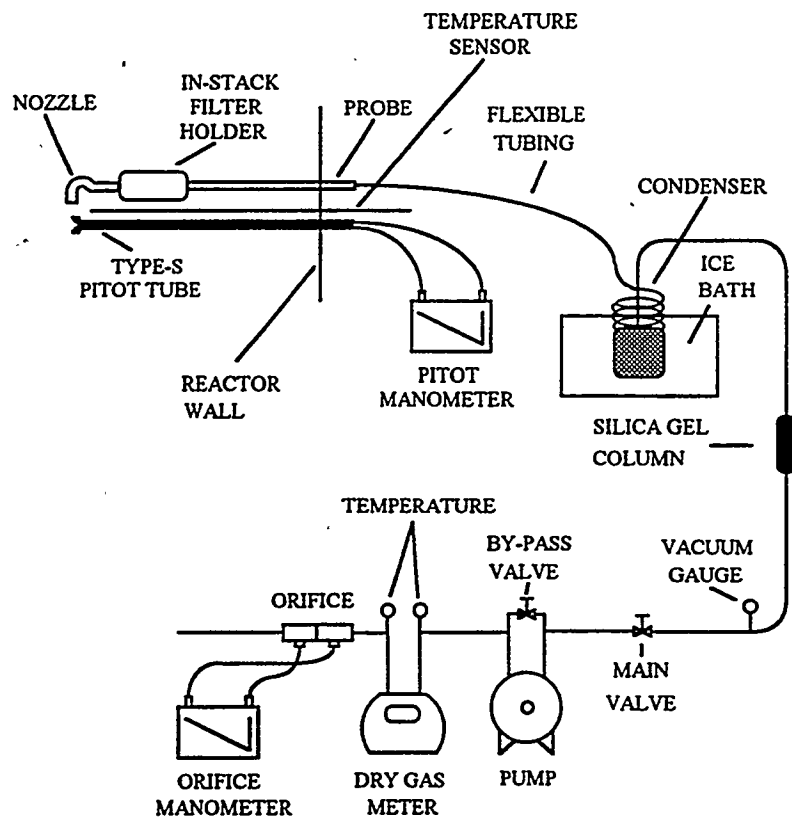
Figure 2-5. Schematic drawing of the mass concentration sampling train.

This train consists of a filter holder and nozzle that is designed to operate in the flue gas stream, followed by a heated sampling probe, a condenser, a drying column, a gas meter, a pump, and finally a flow control mechanism, usually a calibrated orifice. An S-type pitot and thermocouple located near the nozzle provide a means for sampling isokinetically during each test. The glass fiber thimbles are desiccated before and after sampling and then weighed on an electronic microbalance.

Particulate mass concentrations were measured simultaneously upstream of catalyst layer 4 (air heater inlet) and at each air heater outlet (except Reactor C). At the air heater inlet, the three sample ports immediately above the access door at the fourth catalyst layer were used to obtain the sample. During each test particulate was sampled at three, equally-spaced, positions within the three test ports (a total of nine traverse points). At the air heater outlets on reactors A and B, three ports are available for mass sampling. During each test particulate was sampled at

2-6

three, equally-spaced, positions within the three test ports (a total of nine traverse points). Three runs were made at each test site to give a meaningful statistical average.

ASH MINERALOGY, PARTICLE SIZE, ASH RESISTIVITY

After the particulate catches were weighed, they were transported to SRI's Birmingham laboratories where particle size distributions, ash mineralogy and ash resistivity were determined. The instrument used to determine particle size distribution of the fly ash was a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer. It is able to size the particles into approximately 25 size intervals between 0.056 and 56.2 micrometers physical diameter (or Stokes diameter, based on assumed spherical shape and true, or actual, particle density). Ash resistivity was determined using the IEEE 548 (1984) ash resistivity test method, commonly referred to as a descending temperature method. The humidity of the atmosphere surrounding the resistivity test cell was controlled to a moisture content of 8.7%, approximately the same as that measured in the flue gases at the Plant Crist SCR Test Facility.

SAMPLING QUALITY ASSURANCE

The measures adopted to ensure that meaningful results were obtained during the various testing procedures can be divided into three categories; equipment maintenance and calibrations, operating techniques, and analytical techniques. New equipment was obtained for use in the SCR testing program and preventive maintenance and calibrations are performed at regular intervals. Due to the dynamic nature of the process, each measurement requires three replicate samples. The replicates are averaged to yield a representative value. Variability gives a means to discern any anomalies not revealed by other quality control checks.

The sample trains are leak-checked by drawing a vacuum of 15 in. Hg before and after each sampling run. No in-leakage is tolerated. If any loss of pressure is observed within one minute, the source of the leakage is found and eliminated prior to testing. During each run the oxygen level of the gas exiting the dry gas meter is measured to detect air in-leakage that could dilute the samples.

Ammonia samples are analyzed with an ion-specific electrode. The instrument is calibrated using 0.1 µg/mL, 0.5 µg/mL, 1.0 µg/mL, 5.0 µg/mL and 10 µg/mL standards. At the conclusion of each analytical session, a sample spiked with a known amount of ammonia is analyzed and the result compared with the predicted concentration. A blank sample is analyzed to detect zero drift and a mid-range standard is also analyzed to detect calibration drift.

Both the sulfate (SO_2 & SO_3) samples and the chloride (HCl) samples are analyzed by means of ion chromatography. This instrument is calibrated using a minimum of three points on the calibration curve. As with the ammonia analysis, spiked samples, blanks, and standards are analyzed to ensure that calibration drift has not occurred.

While gas-phase flue gas constituents are being sampled, both field blanks and sample blanks are run as quality control checks. The field blank consists of a container of type 1 de-ionized water that is exposed to ambient air at the sampling site. Sample blanks are obtained by passing ambient

air through the probe and through the impingers filled with the appropriate solution. These blanks would detect sample contamination should it occur.

Section 3

AIR HEATER TEST RESULTS

AMMONIA CONCENTRATIONS

Concentrations of ammonia were measured simultaneously at each of the three air heater inlet and outlet test locations. The tests were conducted so as to differentiate between gas-phase and solid-phase ammonia. All three air heaters were tested with the reactors operating at Test Condition 22, the baseline operating condition. The Reactor B and C air heaters were also tested at Test Condition 24 ($NH_3/NO_x = 1.0$), however, the Reactor A air heater was not tested at this condition because of concern that the air heater might become plugged during operation with the higher inlet ammonia concentration.

The test data are summarized in Table 3-1. Table 3-1 also gives estimated solid-phase ammonia concentrations on a mass of ammonia per mass of ash basis. The values shown in the table were derived both from solid-phase ammonia measurements made at the same time the gas-phase measurements were made and mass concentration measurements made earlier at the same location. The ammonia partitioning between the gas and solid phases shown in the table are similar for the air heaters on reactors B and C. These data show that ammonia partitioning at the inlet to the air heater is roughly equal (on a flue gas volumetric basis or ammonia mass basis) between the gas and solid phases on reactors B and C. On both reactors the partitioning shifts toward the solid phase at the air heater outlet. As expected, a significant increase in total ammonia concentration was measured during operation at the higher NH_3/NO_x ratio at Test Condition 24 for both the air heater inlet and air heater outlet on Reactors B and C. The test run on the Reactor A air heater at Test Condition 22 indicated a higher total ammonia concentration at the air heater inlet than at the air heater outlet with the ammonia strongly partitioned to the solid phase. The total ammonia measured at the Reactor A air heater outlet was less than one-half of the total ammonia measured at the air heater inlet with the ammonia more evenly partitioned between the gas and solid phases. The lack of ammonia mass balance closure suggests an error in the measurement so that this result should be disregarded until further testing is done. The original data summaries for these tests can be found in Table A-1 in Appendix A.

SO$_2$ and SO$_3$ CONCENTRATIONS

Concentrations of sulfur dioxide and sulfur trioxide were measured simultaneously at the air heater inlet and outlet test locations of Reactors A, B, and C during operation at Test Condition 22. The test data are summarized in Table 3-2.

SO_3 inlet concentrations ranged from 8 to 19 ppm(v) dry @ 3% O_2, while SO_3 concentrations at the reactor outlets ranged from 5 to 11 ppm(v) dry @ 3% O_2. Reductions in SO_3 concentration across the air heaters ranged from 37.5% on Reactor B to 42.1% on Reactor A. Within the tolerance of the standard deviations given for average SO_2 concentrations in Table 3-2, there was no measurable change in SO_2 concentration across any of the three air heaters. Air heater inlet SO_2 concentrations averaged 2044 ± 41 ppm(v) dry @ 3% O_2, while the air heater outlet SO_2 concentrations averaged 2095 ± 34 ppm(v) dry @ 3% O_2. Since the standard deviations of the mean concentrations at the inlet and outlet overlap, it can be concluded that no measurable

Table 3-1. SCR Reactor Air Heater Tests: Ammonia Concentrations

Reactor	Date	$\frac{NH_3}{NO_x}$	Air Preheater Location	Gas-Phase NH_3, ppm(v), @ 3% O_2, dry	Solid-Phase NH_3, Equivalent ppm(v), @ 3% O_2, dry	Estimated Solid-Phase NH_3, µg/g	Estimated Upper Limit Solid-Phase NH_3, µg/g	Estimated Lower Limit Solid-Phase NH_3, µg/g
A	11-May-94	0.8	Inlet	0.6 ± 0.1	3.3 ± 0.2	307	372	256
			Outlet	0.6 ± 0.1	1.0 ± 0.2	108	139	84
A	Not Tested	1.0	Inlet					
			Outlet					
B	10-May-94	0.8	Inlet	< 0.4	0.2 ± 0.01	22	24	20
			Outlet	<0.3	0.5 ± 0.01	65	70	61
B	10-May-94	1.0	Inlet	1.8 ± 0.4	1.9 ± 0.5	186	239	137
			Outlet	< 0.4	1.3 ± 0.2	162	189	137
C	6-May-94	0.8	Inlet	0.7 ± 0.1	0.6 ± 0.1	56	68	45
			Outlet	< 0.3	1.0 ± 0.1	116	137	99
C	9-May-94	1.0	Inlet	3.3 ± 0.3	3.8 ± 0.5	332	390	280
			Outlet	< 0.3	6.7 ± 1.6	746	1003	525

Table 3-2. SCR Reactor Air Heater Tests: SO_2 and SO_3 Concentrations

| Reactor | SO_3 | | SO_2 | |
| | Inlet | Outlet | Inlet | Outlet |
	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2
A	19 ± 1.4	11 ± 1.2	2040 ± 23	2073 ± 19
B	8.0 ± 1.5	5.0 ± 0.4	2087 ± 17	2078 ± 15
C	17 ± 0.5	10 ± 3.2	2005 ± 50	2134 ± 144

difference in SO_2 concentration was observed. Table A-2 in Appendix A contains the original data summaries for these tests.

HCl CONCENTRATION

Measurements of the concentration of hydrogen chloride were conducted simultaneously at the inlet and outlet of each of the three large reactor air heaters. Three independent tests were performed at each test location. Single point sampling was used. The test results are presented in Table 3-3 as the average concentration and the standard deviation. All HCl concentrations fell within the range of 221 to 233 ppm(v) dry @ 3% O_2. There was no measurable change in HCl concentration across any of the three air heaters. The original data summaries for these HCl tests are presented in Table A-3 in Appendix A.

PARTICULATE MASS CONCENTRATION

Particulate mass concentrations were measured simultaneously at the inlet and outlet of the air heaters on Reactors A and B. Particulate mass concentration was measured only at the inlet to the heat pipe air heater on Reactor C. No outlet measurement was possible on the Reactor C air heater because sampling ports in a location suitable for mass concentration measurements could not be installed on that reactor.

Table 3-4 presents the mass concentration data for the air heater tests. The average air heater inlet mass concentration ranged from 3.42 ± 0.12 (Reactor B) to 3.84 ± 0.17 gr/dscf (Reactor C). The Reactor A inlet and outlet average mass concentration values were within one standard deviation of each other, while the Reactor B average outlet mass concentration was 19% lower than the Reactor B average inlet mass concentration. The original data summaries for the mass concentration tests can be found in Table A-4 in Appendix A.

PARTICLE SIZE DISTRIBUTIONS

The particulate collected during the air heater inlet and outlet particulate mass concentration tests was analyzed using a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer to determine particle size distribution. Table 3-5 presents the mass median diameter (MMD) as the Stokes' diameter (micrometers) for each of the ash samples. The Stokes' or physical diameter is based on the assumption of spherical particles and the true, or actual, particle density. The mass median diameters ranged from 9.9 to 10.4 micrometers at the air heater inlets, and were 10.5 and 11.2 micrometers at the outlet of the Reactor A and B air heaters, respectively. No significant change in MMD across the air heaters was evident from these data. Figures A-1 through A-5 in Appendix A show the particle size distributions graphically on both a cumulative per cent mass basis and a differential mass basis. The figures visually demonstrate the similarity between the particle size distributions among the various ash samples.

ASH MINERALOGY

Fly ash samples collected during mass concentration tests at the air heater inlet and outlet test locations were submitted for ash mineralogy tests. Fly ash samples were available for air heater

Table 3-3. SCR Reactor Air Heater Tests: HCl Concentration

Reactor	HCl	
	Inlet	Outlet
	ppm(v), dry @ 3% O_2	ppm(v), dry @ 3% O_2
A	233 ± 9.8	227 ± 11
B	222 ± 5.7	221 ± 7.2
C	233 ± 2.8	221 ± 8.5

Table 3-4. SCR Reactor Air Heater Tests: Inlet and Outlet Mass Concentrations

Reactor	Air Heater Inlet gr/dscf	Air Heater Outlet gr/dscf
A	3.52 ± 0.44	3.21 ± 0.32
B	3.42 ± 0.12	2.76 ± 0.14
C	3.84 ± 0.17	Not Tested

Table 3-5. SCR Reactor Air Heater Tests: Fly Ash Particle Size

Reactor	Air Heater Inlet Stokes' MMD, micrometers	Air Heater Outlet Stokes' MMD, micrometers
A	9.9	11.2
B	10.4	10.5
C	10.0	Not Tested

inlets on Reactors A, B and C and for the air heater outlets on Reactors A and B. No fly ash samples were collected at the outlet of the Reactor C air heater. The test results are presented in Table 3-6. There were no significant differences in the chemical constituents of the fly ash between the air heater inlets and outlets or among the three air heaters.

ASH RESISTIVITY

Laboratory measurements of ash resistivity were conducted on fly ash samples collected at the inlet and outlet of the three large reactor air heaters, except the outlet of the Reactor C air heater. The test data (pages A-33 to A-38) are summarized in Figure 3-1. The resistivity/temperature relationships for the five ash samples are very similar. For all of the ash samples the peak in the resistivity ranging from 4.5×10^{11} to 1×10^{12} ohm-cm, occurs at about 300 °F. The ash resistivity measured during the Task 1 baseline testing at the location of the SCR test facility inlet scoop was 3.9×10^{11} ohm-cm @ 293 °F. Based on these two measurements, the SCR process appears to have little or no effect on fly ash resistivity.

Table 3-6. SCR Reactor Air Heater Tests: Ash Mineralogy

Chemical Constituent	Reactor A Inlet % weight	Reactor A Outlet % weight	Reactor B Inlet % weight	Reactor B Outlet % weight	Reactor C Inlet % weight
Li_2O	0.03	0.03	0.03	0.03	0.04
Na_2O	0.97	0.91	1.2	1.1	0.91
K_2O	2.4	2.2	2.4	2.5	2.5
MgO	0.98	0.97	1.0	0.97	1.0
CaO	3.5	3.7	2.9	2.9	3.3
Fe_2O_3	18.7	20.3	18.5	18.2	18.5
Al_2O_3	22.2	22.0	22.3	21.7	21.9
SiO_2	48.0	47.4	49.3	49.3	49.0
TiO_2	1.1	1.1	1.1	0.91	0.91
P_2O_5	0.14	0.14	0.14	0.15	0.14
SO_3	1.1	1.0	0.78	0.27	0.78
Loss on Ignition	14.7	16.5	16.3	16.9	16.3

LABORATORY DUST RESISTIVITY
PLANT CRIST SCR TEST FACILITY
LARGE REACTOR AIR HEATERS
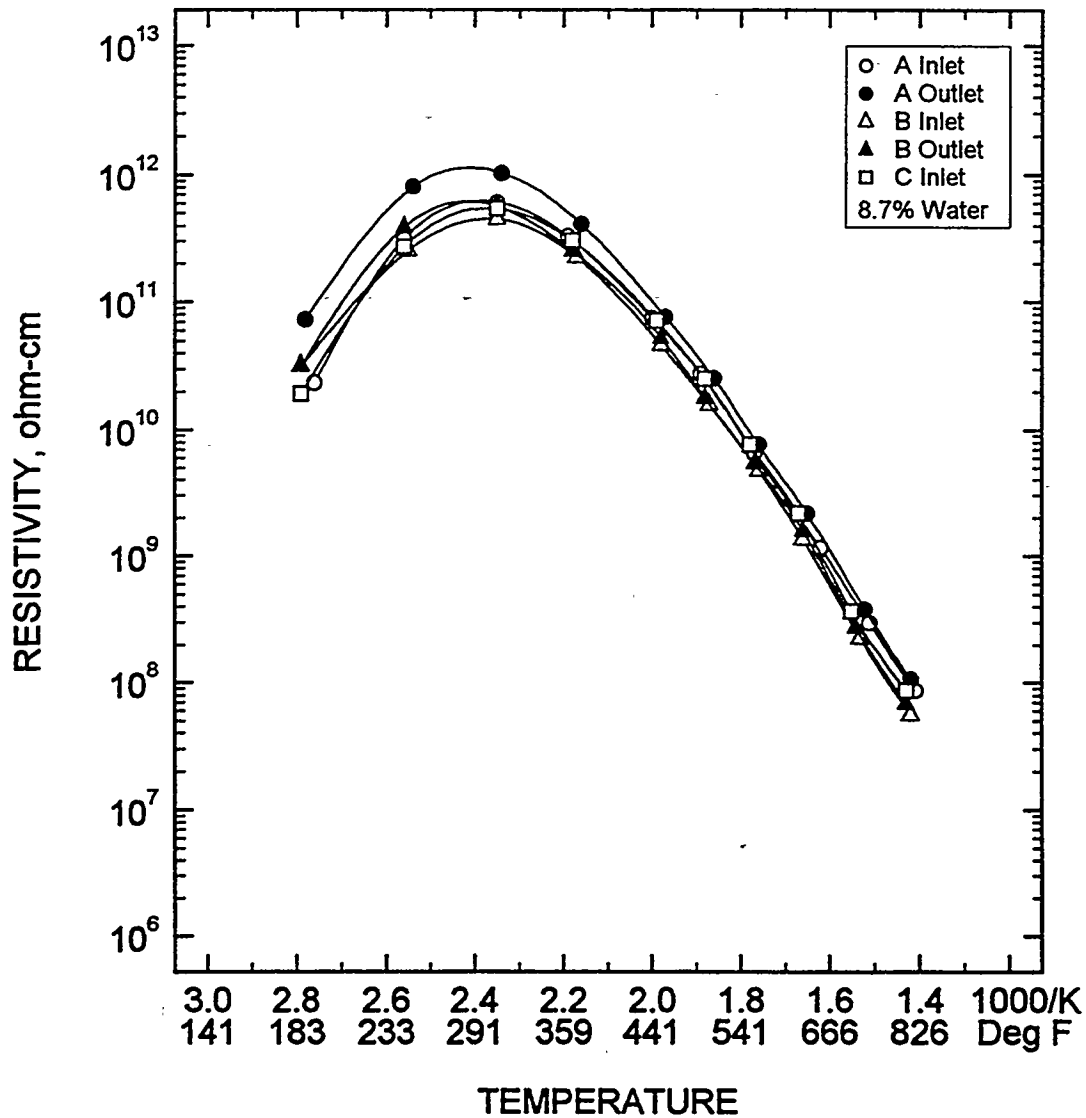
Figure 3-1. Resistivity/temperature relationship for air heater inlet and outlet fly ashes.

Section 4

SUMMARY

This report describes the initial air heater performance characterization tests for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The air heater evaluations were conducted during the last week of April and the first two weeks of May, 1994, following the completion of the first sequence of parametric testing under Task 4: Long-Term Parametric Tests.

The SCR test facility at the Plant Crist test site includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 ^{o}F down to 300 ^{o}F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C incorporates a heat pipe design.

The air heater testing included the determinations of particulate mass concentration and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. Tests were performed at the inlet and outlet of each of the three air heaters (except for mass concentration measurements at the outlet of the Reactor C air heater). The fly ash (particulate) catches from the mass concentration measurements were further analyzed in the laboratory to determine particle size distributions, ash mineralogy, and ash resistivity.

AMMONIA CONCENTRATIONS

Concentrations of ammonia were measured simultaneously at each of the three air heater inlet and outlet test locations. The tests were conducted so as to differentiate between gas-phase and solid-phase ammonia. The majority of the ammonia was found in the solid phase. Most of the gas phase ammonia concentrations were below the detection limit, especially at the air heater outlets. A measurable increase in ammonia concentration in the solid phase was detected during operation at the higher NH_3/NO_x ratio during operation at Test Condition 24 (1.8 to 6.7 ppm(v) dry @ 3% O_2 at Test Condition 24 versus 0.2 to 1.0 ppm(v) dry @ 3% O_2 at Test Condition 22 on Reactors B and C).

SO$_2$ and SO$_3$ CONCENTRATIONS

Concentrations of sulfur dioxide and sulfur trioxide were measured simultaneously at the air heater inlet and outlet test locations of Reactors A, B, and C during operation at Test Condition 22. SO_3 inlet concentrations ranged from 8 to 19 ppm(v) dry @ 3% O_2, while SO_3 concentrations at the reactor outlets ranged from 5 to 11 ppm(v) dry @ 3% O_2. Reductions in SO_3 concentration across the air heaters ranged from 37.5% on Reactor B to 42.1% on Reactor A. Within the

tolerance of the standard deviations for the average SO_2 concentrations, there was no measurable change in SO_2 concentration across any of the three air heaters. Air heater inlet SO_2 concentrations averaged 2044 ± 41 ppm(v) dry @ 3% O_2, while the air heater outlet SO_2 concentrations averaged 2095 ± 34 ppm(v) dry @ 3% O_2.

HCl CONCENTRATION

Measurements of the concentration of hydrogen chloride were conducted simultaneously at the inlet and outlet of each of the three large reactor air heaters. All HCl concentrations fell within the range of 221 to 233 ppm(v) dry @ 3% O_2. There was no measurable change in HCl concentration across any of the three air heaters.

MASS CONCENTRATION

Mass concentration was measured simultaneously at the inlet and outlet of the air heaters on Reactors A and B. Mass concentration was measured only at the inlet to the heat pipe air heater on Reactor C.

The average air heater inlet mass concentrations ranged from 3.42 ± 0.12 (Reactor B) to 3.84 ± 0.17 gr/dscf (Reactor C). The Reactor A inlet and outlet average mass concentration values were within one standard deviation of each other, while the Reactor B average outlet mass concentration was 19% lower than the Reactor B average inlet mass concentration.

PARTICLE SIZE DISTRIBUTIONS

The particulate collected in the Method 17 thimbles during the reactor inlet and outlet mass concentration tests was analyzed using a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer to determine particle size distribution. The mass median diameters (Stokes' diameters) of the fly ashes ranged from 9.9 to 11.2 micrometers. The Stokes' or physical diameter is based on the assumption of spherical particles and the true, or actual, particle density. No significant change in MMD across the air heaters was evident in these data.

ASH MINERALOGY

Fly ash samples collected during mass concentration tests at the air heater inlet and outlet test locations were submitted for ash mineralogy tests. Fly ash samples were available for air heater inlets on Reactors A, B and C and for the air heater outlets on Reactors A and B. No fly ash samples were collected at the outlet of the Reactor C air heater. There were no significant differences in the chemical constituents of the fly ash between the air heater inlets and outlets or among the three air heaters.

ASH RESISTIVITY

Laboratory measurements of ash resistivity were conducted on fly ash samples collected at the inlet and outlet of the three large reactor air heaters, except the outlet of the Reactor C air heater. The

resistivity/temperature relationships for the five ash samples are very similar. For all of the ash samples the peak in the resistivity, ranging from 4.5 x 10^{11} to 1 x 10^{12} ohm-cm, occurs at about 300 ^{o}F. These ash resistivities are similar to those measured during Task 1 on ashes collected from the Unit 5 ESP inlet duct.

APPENDIX A

ORIGINAL DATA SUMMARIES

Table A-1 (Reference Table 3-1)

Work	Source	Run	Time, Begin	Time, End	Date	Avg	±	σ	Cond. No.	Flow lscfm	Temp deg F	NH3/NOx	H2O %	Load MW	Sample O2 %	NH3 mg/l	Df	Vol. ml	NH3 µM	NH3 ppm wet @ 3% O2	Vol. Total STD L wet	Temp Avg deg K	Baro. in.Hg	Temp Begin deg F	Temp End deg F	Vol. Begin ft3	Vol. End ft3	Meter I.D.	Corr. Factor
071	AAI NHx	01	0930	0941	11-May	<0.7	±	BDL	Air Blank							0.100 / 0.100	1	343	2 / 2	0.70 / 0.70	77.92	302	30.13	83	86	425.100	428.100	Stack 1	1.0080
071	AAI NHQ	02	0950	1033	11-May	0.56	±	0.03	22	5	700	0.8	9	82	3.8	0.156 / 0.168	1	351	4 / 4	0.54 / 0.58	169.38	306	30.13	86	95	428.200	434.200	Stack 1	1.0080
071	AAI NHQ	03	1050	1133	11-May	0.51	±	0.02	22	5	700	0.8	9	82	3.8	0.151 / 0.143		349	4 / 4	0.52 / 0.49	168.15	308	30.13	94	95	434.400	440.400	Stack 1	1.0080
071	AAI NHQ	04	1146	1233	11-May	0.73	±	0.00	22	5	700	0.8	9	83	4.3	0.223 / 0.224		319	5 / 5	0.73 / 0.73	168.31	307	30.13	95	93	440.600	446.600	Stack 1	1.0080
						0.6	±	0.12							3.8 / 0.4														
071	AAI NHS	02	0930	1033	11-May	3.12	±	0.06	22	5	700	0.8	9	82	3.8	1.430 / 1.470	1	220	22 / 23	3.07 / 3.16	169.38	306	30.13	86	95	428.200	434.200	Stack 1	1.0080
071	AAI NHS	03	1050	1133	11-May	0.50	±	0.04	22	5	700	0.8	9	83	3.8	0.249 / 0.238		225	4 / 4	0.49 / 0.54	168.15	308	30.13	94	95	434.400	440.400	Stack 1	1.0080
071	AAI NHS	04	1146	1233	11-May	3.39	±	0.08	22	5	700	0.8	9	83	4.3	1.830 / 1.890		178	23 / 24	3.34 / 3.45	168.31	307	30.13	95	93	440.600	446.600	Stack 1	1.0080
						3.3	±	0.20							3.8 / 0.4														
071	AAO NHx	01	0925	0941	11-May	<0.4	±	BDL	Air Blank							0.100 / 0.100	1	204	1 / 1	0.43 / 0.43	76.01	306	30.13	90	93	653.500	656.500	Stack 2	0.9960
071	AAO NHQ	02	1000	1040	11-May	0.29	±	0.00	22	5	700	0.8	9	82	7.5	0.100 / 0.100	1	227	2 / 2	0.29 / 0.29	166.30	307	30.13	92	96	656.700	662.700	Stack 2	0.9960
071	AAO NHQ	03	1100	1137	11-May	0.28	±	0.00	22	5	700	0.8	9	82	7.4	0.100 / 0.100		218	2 / 2	0.28 / 0.28	165.55	309	30.13	95	98	662.900	668.900	Stack 2	0.9960
071	AAO NHQ	04	1155	1235	11-May	0.36	±	0.00	22	5	700	0.8	9	83	7.8	0.100 / 0.100		273	2 / 2	0.36 / 0.36	165.70	309	30.13	97	95	669.000	675.000	Stack 2	0.9960
						<0.3	±	BDL							7.6 / 0.2														
071	AAO NHS	02	1000	1040	11-May	1.15	±	0.04	22	5	700	0.8	9	82	7.5	0.520 / 0.544	1	168	6 / 7	1.12 / 1.17	166.30	307	30.13	92	96	656.700	662.700	Stack 2	0.9960
071	AAO NHS	03	1100	1137	11-May	1.13	±	0.03	22	5	700	0.8	9	82	7.4	0.705 / 0.732		123	6 / 6	1.11 / 1.15	165.55	309	30.13	95	98	662.900	668.900	Stack 2	0.9960
071	AAO NHS	04	1155	1235	11-May	0.86	±	0.02	22	5	700	0.8	9	83	7.8	0.466 / 0.483		138	5 / 5	0.85 / 0.88	165.70	309	30.13	97	95	669.000	675.000	Stack 2	0.9960
						1.0	±	0.16							7.6 / 0.2														

A-2

Table A-1, continued (Reference Table 3-1)

				Concentration			Reactor Conditions							Analytical					Dry Gas Meter								
W S G N	Time, Begin	Time, End	Date	ppm (V) wet @ 3% O2 Average	±	σ	Condition Number	Flow kscfm	Temp. degF	NH3/NOx	H2O %	Unit 5 Load, MW	Sample O2 %	NH3 mg/l	Df	Vol. ml	NH3 µM	NH3 ppm wet @ 3% O2	Volume, Total STD L, wet	Temp, Avg. degK	Barometric Pressure In. Hg	Temp, Begin degF	Temp, End degF	Volume, Begin ft3	Volume, End ft3	Meter I.D	Corr. Factor
071 BA1 NHx 01	0914	0926	10-May	<0.56	±	0.00	Air Blank							0.100 / 0.100	1	277	2.0 / 2.0	0.56 / 0.56	79.12	297	30.09	76	75	365.700	368.700	Stack 1	1.0080
071 BA1 NHQ 02	0941	1030	10-May	0.31	±	0.00	22	5	700	0.8	9	77	4.7	0.100 / 0.100	1	307	2.2 / 2.2	0.31 / 0.31	173.41	298	30.09	76	78	369.000	375.000	Stack 1	1.0080
071 BA1 NHQ 03	1054	1142	10-May	0.36	±	0.00	22	5	700	0.8	9	83	2.8	0.100 / 0.100	1	390	2.8 / 2.8	0.36 / 0.36	172.44	300	30.09	79	81	375.200	381.200	Stack 1	1.0080
071 BA1 NHQ 04	1200	1248	10-May	0.42	±	0.00	22	5	700	0.8	9	80	3.5	0.124 / 0.124	1	356	3.2 / 3.2	0.42 / 0.42	172.28	300	30.09	80	81	381.400	387.400	Stack 1	1.0080
				<0.4	±	0.05							3.7 / 1.0					0.1									
071 BA1 NHS 02	0941	1030	10-May	0.24	±	0.01	22	5	700	0.8	9	77	4.7	0.110 / 0.104	1	217	1.7 / 1.6	0.24 / 0.23	173.41	298	30.09	76	78	369.000	375.000	Stack 1	1.0080
071 BA1 NHS 03	1054	1142	10-May	0.23	±	0.00	22	5	700	0.8	9	83	2.8	0.105 / 0.104	1	238	1.8 / 1.8	0.23 / 0.23	172.44	300	30.09	79	81	375.200	381.200	Stack 1	1.0080
071 BA1 NHS 04	1200	1248	10-May	0.21	±	0.01	22	5	700	0.8	9	80	3.5	0.101 / 0.097	1	226	1.6 / 1.6	0.22 / 0.21	172.28	300	30.09	80	81	381.400	387.400	Stack 1	1.0080
				0.2	±	0.01							3.7 / 1.0					0.0									
071 BAO NHx 01	0913	0926	10-May	<0.3	±	0.00	Air Blank							0.100 / 0.100	1	158	1.1 / 1.1	0.33 / 0.33	77.53	300	30.09	80	80	612.700	615.700	Stack 2	0.9960
071 BAO NHQ 02	0950	1025	10-May	0.28	±	0.00	22	5	700	0.8	9	77	7.5	0.100 / 0.100	1	219	1.6 / 1.6	0.28 / 0.28	169.92	301	30.09	80	83	615.900	621.900	Stack 2	0.9960
071 BAO NHQ 03	1105	1137	10-May	0.31	±	0.00	22	5	700	0.8	9	82	7.2	0.100 / 0.100	1	248	1.8 / 1.8	0.31 / 0.31	169.29	302	30.09	84	83	622.100	628.100	Stack 2	0.9960
071 BAO NHQ 04	1207	1240	10-May	0.31	±	0.00	22	5	700	0.8	9	80	7.5	0.100 / 0.100	1	244	1.7 / 1.7	0.31 / 0.31	169.45	301	30.09	82	84	628.300	634.300	Stack 2	0.9960
				<0.3	±	0.02							7.4 / 0.2					0.0									
071 BAO NHS 02	0950	1025	10-May	0.55	±	0.02	22	5	700	0.8	9	77	7.5	0.267 / 0.281	1	160	3.1 / 3.2	0.54 / 0.57	169.92	301	30.09	80	83	615.900	621.900	Stack 2	0.9960
071 BAO NHS 03	1105	1137	10-May	0.53	±	0.01	22	5	700	0.8	9	82	7.2	0.211 / 0.216	1	201	3.0 / 3.1	0.52 / 0.54	169.29	302	30.09	84	83	622.100	628.100	Stack 2	0.9960
071 BAO NHS 04	1207	1240	10-May	0.54	±	0.00	22	5	700	0.8	9	80	7.5	0.272 / 0.273	1	157	3.1 / 3.1	0.54 / 0.54	169.45	301	30.09	82	84	628.300	634.300	Stack 2	0.9960
				0.5	±	0.01							7.4 / 0.2					0.0									

Table A-1, continued (Reference Table 3-1)

							Concentration			Reactor Conditions							Analytical					Dry Gas Meter								
				Time, Begin	Time, End	Date	ppm (v) wet @ 3% O2 Average	±	σ	Condition Number	Flow kscfm	Temp. deg F	NH3/NOx	H2O %	Unit 5 Load MW	Sample O2 %	NH3 mg/l	Df	Vol. ml	NH3 µM	NH3 ppm wet @ 3% O2	Volume Total STD L, wet	Temp Avg deg K	Barometric Pressure in. Hg	Temp Begin deg F	Temp End deg F	Volume Begin ft3	Volume End ft3	Meter I.D.	Corr. Factor
071	BAI	NHx	01	0914	0926	10-May	<0.56	±	0.00	Air Blank																				
071	BAI	NHQ	05	1435	1520	10-May	1.35	±	0.00	24	5	700	1	9	81	3.6	0.358	1	391	10.0	1.36	170.86	302	30.09	83	87	388.000	394.000	Stack 1	1.0080
																	0.357			10.0	1.35									
071	BAI	NHQ	06	1543	1627	10-May	2.24	±	0.05	24	5	700	1	9	83	4.0	0.622		368	16.3	2.28	170.23	304	30.09	87	87	394.200	400.200	Stack 1	1.0080
																	0.603			15.9	2.21									
071	BAI	NHQ	07	1644	1731	10-May	1.80	±	0.08	24	5	700	1	9	83	3.3	0.444		410	13.0	1.75	169.46	305	30.09	88	91	400.400	406.400	Stack 1	1.0080
																	0.472			13.8	1.86									
							1.8	±	0.44							3.6					0.4									
																0.4														
071	BAI	NHS	05	1442	1516	10-May	1.40	±	0.01	24	5	700	1	9	79	7.5	0.496	1	222	7.9	1.40	168.21	304	30.09	86	88	634.700	640.700	Stack 2	0.9960
																	0.500			7.9	1.41									
071	BAI	NHS	06	1547	1621	10-May	2.08	±	0.03	24	5	700	1	9	82	7.2	0.758		218	11.8	2.06	167.59	305	30.09	88	90	640.900	646.900	Stack 2	0.9960
																	0.773			12.0	2.10									
071	BAI	NHS	07	1653	1728	10-May	2.27	±	0.05	24	5	700	1	9	83	7.0	1.010		180	13.0	2.24	167.14	306	30.09	90	91	647.200	653.200	Stack 2	0.9960
																	1.040			13.4	2.31									
							1.9	±	0.46							7.2					0.5									
																0.3														
071	BAO	NHx	01	0913	0926	10-May	<0.3	±	0.00	Air Blank																				
070	BAO	NHQ	05	1442	1516	10-May	0.30	±	0.00	24	5	700	1	9	79	7.5	0.100	1	240	1.7	0.30	168.21	304	30.09	86	88	634.700	640.700	Stack 2	0.9960
																	0.100			1.7	0.30									
070	BAO	NHQ	06	1547	1621	10-May	0.33	±	0.00	24	5	700	1	9	82	7.2	0.100		266	1.9	0.33	167.59	305	30.09	88	90	640.900	646.900	Stack 2	0.9960
																	0.100			1.9	0.33									
070	BAO	NHQ	07	1653	1728	10-May	0.43	±	0.00	24	5	700	1	9	83	7.0	0.100		349	2.5	0.43	167.14	306	30.09	90	91	647.200	653.200	Stack 2	0.9960
																	0.100			2.5	0.43									
							<0.4	±	0.1							7.2					0.1									
																0.3														
070	BAO	NHS	05	1442	1516	10-May	1.18	±	0.05	24	5	700	1	9	79	7.5	0.536	1	167	6.4	1.14	168.21	304	30.09	86	88	634.700	640.700	Stack 2	0.9960
																	0.572			6.8	1.21									
070	BAO	NHS	06	1547	1621	10-May	1.47	±	0.00	24	5	700	1	9	82	7.2	1.180		99.6	8.4	1.47	167.59	305	30.09	88	90	640.900	646.900	Stack 2	0.9960
																	1.179			8.4	1.46									
070	BAO	NHS	07	1653	1728	10-May	1.39	±	0.02	24	5	700	1	9	83	7.0	0.744		150	8.0	1.38	167.14	306	30.09	90	91	647.200	653.200	Stack 2	0.9960
																	0.761			8.2	1.41									
							1.3	±	0.2							7.2					0.2									
																0.3														

Table A-1, continued (Reference Table 3-1)

Wk	Src	Run	Time Begin	Time End	Date	Conc. Avg	±	σ	Cond. No.	Flow kscfm	Temp degF	NH3/NOx	H2O %	Load MW	Sample O2 %	NH3 mg/l	Df	Vol ml	NH3 µM	NH3 ppm wet @3% O2	Vol Total STD L wet	Temp Avg degK	Baro in.Hg	Temp Begin degF	Temp End degF	Vol Begin ft3	Vol End ft3	Meter I.D	Corr. Factor
070	CAI	NHx 01	1005	1020	6-May	<0.3	±	0.00	Air Blank							0.100	1	188	1	0.30	99.04	302	30.18	82	85	302.500	306.300	Stack 1	1.0080
																0.100			1	0.30									
070	CAI	NHG 02	1104	1155	6-May	0.77	±	0.00	22	5	700	0.8	9	84	3.1	0.259	1	313	6	0.77	170.12	305	30.18	86	92	306.500	312.500	Stack 1	1.0080
																0.261			6	0.77									
070	CAI	NHG 03	1341	1428	6-May	0.76	±	0.01	22	5	700	0.8	9	85	2.9	0.254		317	6	0.76	169.35	306	30.18	90	93	312.700	318.700	Stack 1	1.0080
																0.257			6	0.77									
070	CAI	NHG 04	1455	1543	6-May	0.66	±	0.01	22	5	700	0.8	9	85	3.5	0.193		354	5	0.67	168.74	307	30.18	93	94	319.000	325.000	Stack 1	1.0080
																0.189			5	0.65									
						0.7	±	0.06							3.2					0.1									
															0.3														
070	CAI	NHS 02	1104	1155	6-May	0.77	±	0.01	22	5	700	0.8	9	84	3.1	0.341		236	6	0.76	170.12	305	30.18	86	92	306.500	312.500	Stack 1	1.0080
																0.347			6	0.77									
070	CAI	NHS 03	1341	1428	6-May	0.56	±	0.00	22	5	700	0.8	9	85	2.9	0.291		206	4	0.56	169.35	306	30.18	90	93	312.700	318.700	Stack 1	1.0080
																0.292			4	0.57									
070	CAI	NHS 04	1455	1543	6-May	0.61	±	0.00	22	5	700	0.8	9	85	3.5	0.277		224	4	0.61	168.74	307	30.18	93	94	319.000	325.000	Stack 1	1.0080
																0.279			4	0.61									
						0.6	±	0.11							3.2					0.1									
															0.3														
070	CAO	NHx 01	1024	1037	6-May	<0.4	±	0.00	Air Blank							0.100	1	181	1	0.38	77.05	302	30.18	84	86	568.400	571.400	Stack 2	0.9960
																0.100	1		1	0.38									
070	CAO	NHG 02	1118	1155	6-May	0.29	±	0.00	22	5	700	0.8	9	84	6.0	0.100	1	255	2	0.29	168.40	304	30.18	86	90	572.000	578.000	Stack 2	0.9960
																0.100			2	0.29									
070	CAO	NHG 03	1350	1423	6-May	0.29	±	0.00	22	5	700	0.8	9	85	5.9	0.100		251	2	0.29	167.94	305	30.18	88	91	578.200	584.200	Stack 2	0.9960
																0.100			2	0.29									
070	CAO	NHG 04	1509	1541	6-May	0.24	±	0.00	22	5	700	0.8	9	85	5.6	0.100		229	2	0.24	178.49	306	30.18	91	92	584.400	590.800	Stack 2	0.9960
																0.100			2	0.24									
						<0.3	±	0.03							5.8					0.0									
															0.2														
070	CAO	NHS 02	1118	1155	6-May	1.14	±	0.01	22	5	700	0.8	9	84	6.0	0.403		247	7	1.14	168.40	304	30.18	86	90	572.000	578.000	Stack 2	0.9960
																0.406			7	1.14									
070	CAO	NHS 03	1350	1423	6-May	1.03	±	0.02	22	5	700	0.8	9	85	5.9	0.418		215	6	1.02	167.94	305	30.18	88	91	578.200	584.200	Stack 2	0.9960
																0.428			7	1.05									
070	CAO	NHS 04	1509	1541	6-May	0.96	±	0.00	22	5	700	0.8	9	85	5.6	0.514		178	7	0.96	178.49	306		91	92	584.400	590.800	Stack 2	0.9960
																0.514			7	0.96									
						1.0	±	0.09							5.8					0.1									
															0.2														

Table A-1, continued (Reference Table 3-1)

Week				Time Begin	Time End	Date	Concentration ppm (v), wet, @3% O2 Average	±	σ	Reactor Conditions Condition Number	Flow kscfm	Temp deg F	NH3/NO2	H2O %	Unit 5 Load, MW	Sample O2 %	Analytical NH3 mg/l	Df	Vol. ml	NH3 µM	NH3 ppm, wet, @3% O2	Dry Gas Meter Volume, Total STD L, wet	Temp, Avg. deg K	Barometric Pressure In. Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin ft3	Volume, End ft3	Meter I.D	Corr. Factor
071	CAI	NHx	01	1000	1014	9-May	<.4	±	0.00	Air Blank							0.100	1	206	1	0.43	77.23	304	30.03	88	87	336.300	339.300	Stack 1	1.0080
																	0.100			1	0.43									
071	CAI	NHG	02	1031	1121	9-May	3.09	±	0.00	24	5	700	1	9	86	2.3	1.050	1	323	24	3.09	169.12	305	30.03	88	91	339.600	345.600	Stack 1	1.0080
																	1.051			24	3.09									
071	CAI	NHG	03	1303	1353	9-May	3.57	±	0.02	24	5	700	1	9	86	2.6	1.160		331	27	3.55	169.12	305	30.03	94	85	345.800	351.800	Stack 1	1.0080
																	1.170			28	3.58									
071	CAI	NHG	04	1410	1458	9-May	3.15	±	0.05	24	5	700	1	9	85	3.1	0.915		357	23	3.11	168.97	305	30.03	86	94	352.100	358.100	Stack 1	1.0080
																	0.937			24	3.19									
							3.3	±	0.26							2.7					0.3									
																0.4														
071	CAI	NHS	02	1031	1121	9-May	4.27	±	0.04	24	5	700	1	9	86	2.3	2.070	1	225	33	4.24	169.12	305	30.03	88	91	339.600	345.600	Stack 1	1.0080
																	2.100			34	4.30									
071	CAI	NHS	03	1303	1353	9-May	3.90	±	0.04	24	5	700	1	9	86	2.6	1.890		221	30	3.87	169.12	305	30.03	94	85	345.800	351.800	Stack 1	1.0080
																	1.920			30	3.93									
071	CAI	NHS	04	1410	1458	9-May	3.36	±	0.05	24	5	700	1	9	85	3.1	1.920		182	25	3.33	168.97	305	30.03	86	94	352.100	358.100	Stack 1	1.0080
																	1.960			25	3.40									
							3.8	±	0.46							2.7					0.5									
																0.4														
071	CAO	NHx	01	0957	1010	9-May	<0.3	±	0.00	Air Blank							0.100	1	144	1	0.30	77.09	301	30.03	82	82	591.000	594.000	Stack 2	0.9960
																	0.100			1	0.30									
071	CAO	NHG	02	1045	1115	9-May	0.24	±	0.00	24	5	700	1	9	86	3.6	0.100	1	246	2	0.24	168.49	302	30.03	83	87	594.300	600.300	Stack 2	0.9960
																	0.100			2	0.24									
071	CAO	NHG	03	1316	1351	9-May	0.28	±	0.00	24	5	700	1	9	86	4.5	0.100		264	2	0.28	166.50	306	30.03	90	93	600.200	606.200	Stack 2	0.9960
																	0.100			2	0.28									
071	CAO	NHG	04	1415	1450	9-May	0.24	±	0.00	24	5	700	1	9	85	4.6	0.100		222	2	0.24	165.75	307	30.03	93	95	606.400	612.400	Stack 2	0.9960
																	0.100			2	0.24									
							<0.3	±	0.02							4.2					0.0									
																0.6														
070	CAO	NHS	02	1045	1115	9-May	8.47	±	0.11	24	5	700	1	9	86	3.6	4.770		179	61	8.39	168.49	302	30.03	83	87	594.300	600.300	Stack 2	0.9960
																	4.860			62	8.55									
070	CAO	NHS	03	1316	1351	9-May	6.39	±	0.09	24	5	700	1	9	86	4.5	3.350		180	43	6.32	166.50	306	30.03	90	93	600.200	606.200	Stack 2	0.9960
																	3.420			44	6.46									
070	CAO	NHS	04	1415	1450	9-May	5.26	±	0.05	24	5	700	1	9	85	4.6	2.770		178	35	5.23	165.75	307	30.03	93	95	606.400	612.400	Stack 2	0.9960
																	2.810			36	5.30									
							6.7	±	1.63							4.2					1.6									
																0.6														

Table A-2 (Reference Table 3-2)

				Time, Begin	Time, End	Date	Concentration ppm(s) wet @3%O2 Average	±	σ	Condition Number	Flow (uscfm)	Temp (deg F)	NH3/NOx	H2O %	Unit 5 Boiler Load MW	Sample O2 %	SO4 ppm	Dil	Vol ml	SO4 µM	SO2 or SO3 ppm wet @3%O2	Volume Total STD L wet	Temp Avg deg K	Barometric Pressure in Hg	Temp Begin deg F	Temp End deg F	Volume Begin	Volume End	Meter I.D	Corr. Factor
070	AAI	SO3	01	1050	1105	3-May	17.4	±	0.03	22	5	700	0.8	9	84	2.3	9,180	12.5	58.7	70	17.3	87.21	297	30.01	78	74	235.000	238.020	Stack 1	1.0080
																	9,200			70	17.4									
070	AAI	SO3	02	1143	1157	3-May	19.3	±	0.04	22	5	700	0.8	9	84	2.3	9,705		61.9	78	19.3	87.53	294	30.01	71	70	238.400	241.400	Stack 1	1.0080
																	9,730			78	19.3									
070	AAI	SO3	03	1212	1226	3-May	20.2	±	0.11	22	5	700	0.8	9	84	3.1	6,953		85.8	78	20.1	87.04	296	30.01	72	75	241.800	244.800	Stack 1	1.0080
																	7,005			78	20.3									
																2.6														
							19	±	1.4							0.5														
070	AAI	SO2	01				2031	±	4								6,894	200	573	8230	2034.2									
																	6,874			8206	2028.3									
070	AAI	SO2	02				2022	±	11								6,413		617	8243	2030.1									
																	6,365			8182	2014.9									
070	AAI	SO2	03				2067	±	4								5,644		680	7996	2069.2									
																	5,630			7976	2064.1									
							2040	±	23																					
070	AAO	SO3	01	1050	1104	3-May	10.5	±	0.04	22	5	700	0.8	9	84	5.7	3,616	12.5	70.9	33	10.5	83.73	304	30.01	89	87	507.100	510.100	Stack 2	0.9960
																	3,597			33	10.5									
070	AAO	SO3	02	1142	1155	3-May	4.8	±	0.05	22	6	700	0.8	9	84	6.3	4,464		71.3	45	4.7	84.58	301	30.01	84	84	510.600	513.600	Stack 2	0.9940
																	4,499			45	4.8									
070	AAO	SO3	03	1211	1222	3-May	12.2	±	0.03	22	5	700	0.8	9	84	6.5	3,329		85.7	37	12.2	84.81	300	30.01	81	81	513.900	516.900	Stack 2	0.9960
																	3,317			37	12.2									
																6.2														
							11	±	1.2							0.4														
070	AAO	SO2	01	1050	1104	3-May	2074	±	2								5,299	200	596	6380	2072.9									
																	5,307			6590	2076.0									
070	AAO	SO2	02	1142	1155	3-May	2091	±	27								5,421		565	6581	2071.9									
																	5,522			6500	2110.5									
070	AAO	SO2	03	1211	1222	3-May	2054	±	7								4,245		709	6270	2058.5									
																	4,225			6241	2048.8									
							2073	±	19																					

A-7

Table A-2, continued (Reference Table 3-2)

ID				Time, Begin	Time, End	Date	Average ppm(s),wet,@3%O2	±	σ	Condition Number	Flow kscfm	Temp deg F	NBU/NO2	H2O %	Unit 5 Boiler Load, MW	Sample O2 %	SO4 ppm	Dil	Vol. ml	SO4 µM	SO2, or SO3 ppm,wet @3%O2	Volume, Total STD L, wet	Temp, Avg. deg K	Barometric Pressure in.Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin ft3	Volume, End ft3	Meter I.D	Corr. Factor
070	BAI	SO3	01	1025	1034	4-May	9.5	±	0.10	22	5	700	0.8	9	84	3.5	4.281	12.5	65.7	37	9.6	87.95	293	30.01	67	69	258.000	261.000	Stack 1	1.0080
																	4.217			36	9.5									
070	BAI	SO3	02	1053	1107	4-May	7.7	±	0.01	22	5	700	0.8	9	84	3.6	3.802		59.5	29	7.7	88.28	294	30.01	69	70	261.500	264.520	Stack 1	1.0080
																	3.796			29	7.7									
070	BAI	SO3	03	1155	1209	4-May	6.6	±	0.02	22	5	700	0.8	9	84	3.8	3.057		62.9	25	6.6	87.29	295	30.01	71	73	265.000	268.000	Stack 1	1.0080
																	3.047			25	6.6									
							8.0	±	1.5							3.8														
																0.1														
070	BAI	SO2	01	1025	1034	4-May	2076	±	8								6.218	200	613	7941	2080.6									
																	6.193			7909	2072.2									
070	BAI	SO2	02	1053	1107	4-May	2078	±	4								6.218		612	7928	2081.2									
																	6.201			7906	2075.6									
070	BAI	SO2	03	1155	1209	4-May	2107	±	16								6.052		626	7893	2095.7									
																	6.116			7976	2117.9									
							2087	±	17																					
070	BAO	SO3	01	1025	1038	4-May	4.6	±	0.08	22	5	700	0.8	9	84	6.9	1.155	12.5	93.9	14	4.7	86.09	296	30.01	72	74	528.000	531.000	Stack 2	0.9960
																	1.125			14	4.6									
070	BAO	SO3	02	1053	1106	4-May	5.3	±	0.00	22	5	700	0.8	9	84	6.9	1.382		88.1	16	5.3	85.92	296	30.01	74	74	531.400	534.400	Stack 2	0.9960
																	1.383			16	5.3									
070	BAO	SO3	03	1154	1207	4-May	5.2	±	0.01	22	5	700	0.8	9	84	7.9	1.255		89.2	15	5.2	86.01	296	30.01	73	74	534.700	537.700	Stack 2	0.9960
																	1.258			15	5.2									
							5.1	±	0.4							7.2														
																0.6														
070	BAO	SO2	01	1025	1038	4-May	2088	±	1								5.114	200	589	6275	2087.7									
																	5.117			6279	2088.9									
070	BAO	SO2	02	1053	1106	4-May	2067	±	1								4.818		618	6203	2067.6									
																	4.814			6198	2065.9									
070	BAO	SO2	03	1154	1207	4-May	2245	±	28								4.992		728	6906	2235.6									
																	4.464			6345	2264.8									
							2078	±	15																					

Table A-2, continued (Reference Table 3-2)

				Time Begin	Time End	Date	Conc. ppm (%) wt. @ 3% O2 Average	±	σ	Condition Number	Flow kscfm	Temp deg F	NH3/NO2	H2O %	Unit 5 Boiler Load MW	Sample O2 %	SO4 ppm	Dil	Vol ml	SO4 µM	SO4, or SO3 ppm wt. @ 3% O2	Volume, Total STD L wet	Temp, Avg. deg K	Barometric Pressure in. Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin m3	Volume, End m3	Meter I.D.	Corr. Factor
070	CAI	SO3	01	1047	1101	5-May	16.8	±	0.07	22	5	700	0.8	9	84	4.3	6.924	12.5	67.7	61	16.7	88.08	294	30.14	69	70	281.600	284.600	Stack 1	1.0080
																	6.966			61	16.8									
070	CAI	SO3	02	1116	1130	5-May	17.2	±	0.04	22	5	700	0.8	9	84	4.0	8.074		62.7	66	17.2	90.88	295	30.14	72	72	285.000	288.110	Stack 1	1.0080
																	8.045			66	17.1									
070	CAI	SO3	03	1142	1155	5-May	17.7	±	0.04	22	5	700	0.8	9	84	3.8	8.123		63.1	67	17.7	88.29	296	30.14	73	74	288.500	291.530	Stack 1	1.0080
																	8.099			67	17.7									
							17.2	±	0.5							4.0														
																0.3														
070	CAI	SO2	01	1047	1101	5-May	2062	±	0								5.916	200	609	7506	2058.3									
																	5.939			7535	2066.3									
070	CAI	SO2	01	1116	1130	5-May	1973	±	0								5.990		613	7535	1967.0									
																	5.938			7583	1979.7									
070	CAI	SO2	01	1142	1155	5-May	1979	±	0								5.373		668	7477	1985.7									
																	5.339			7430	1973.2									
							2005	±	50																					
070	CAO	SO3	01	1048	1102	5-May	10.6	±	0.03	22	5	700	0.8	9	84	4.5	3.057	12.5	93.7	37	10.6	86.22	297	30.14	74	75	548.000	551.000	Stack 2	0.9960
																	3.045			37	10.5									
070	CAO	SO3	02	1115	1129	5-May	13.1	±	0.02	22	5	700	0.8	9	84	6.1	3.438		92.5	41	13.0	85.97	297	30.14	75	77	551.700	554.700	Stack 2	0.9960
																	3.444			41	13.1									
070	CAO	SO3	03	1142	1156	5-May	6.7	±	0.05	22	5	700	0.8	9	84	6.1	1.826		88.8	21	6.7	85.73	298	30.14	76	79	555.000	558.000	Stack 2	0.9960
																	1.845			21	6.7									
							10.1	±	3.2							5.8														
																0.9														
070	CAO	SO2	01	1048	1102	5-May	1968.9	±	2.0								4.876	200	684	6948	1970.3									
																	4.869			6938	1967.5									
070	CAO	SO2	02	1115	1129	5-May	2233.0	±	15.0								5.399		633	7120	2243.6									
																	5.348			7053	2222.4									
070	CAO	SO2	03	1142	1156	5-May	2200.9	±	1.3								4.687		713	6962	2200.0									
																	4.691			6968	2201.9									
							2134.3	±	144.1																					

Table A-3 (Reference Table 3-3)

				Concentration ppm (v), wet, @ 3% O₂			Reactor Conditions					Unit 5 Load MW	Sample O₂ %	Analytical					Dry Gas Meter								Corr. Factor
Work/Num	Time, Begin	Time, End	Date	Average	±	σ	Condition Number	Flow uscfm	Temp deg F	NH₃/NOₓ	H₂O %			CT ppm	DI	Vol ml	CT µM	HCl ppm, wet @3% O2	Volume, Total STD L, wet	Temp, Avg. deg K	Barometric Pressure in. Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin ft3	Volume, End ft3	Meter I.D.	
070 AAI HCl 01	1422	1436	3-May	235	±	2.8	22	5	700	0.8	9	84	2.5	4.995	12.5	522	918	233.4	85.68	301	30.01	80	84	245.200	248.200	Stack 1	1.0080
														5.069			933	237.3									
070 AAI HCl 02	1448	1503	3-May	241	±	4.0	22	5	700	0.8	9	84	2.5	4.073	12.5	650	934	238.1	85.44	302	30.01	82	85	248.800	251.800	Stack 1	1.0080
														4.169			956	243.7									
070 AAI HCl 03	1513	1527	3-May	222	±	0.7	22	5	700	0.8	9	84	2.4	3.565	12.5	702	882	221.4	86.37	301	30.01	82	84	252.300	255.330	Stack 1	1.0080
														3.580			886	222.4									
				233	±	8.8						AVG	2.5														
												σ	0.1														
070 AAO HCl 01	1422	1435	3-May	238	±	0.1	22	5	700	0.8	9	84	8.8	1.460	25	688	708	238.2	84.58	301	30.01	82	83	517.460	520.460	Stack 2	0.996
														1.461			709	238.3									
070 AAO HCl 02	1449	1502	3-May	217	±	0.6	22	5	700	0.8	9	84	8.0	1.547	25	622	679	216.7	84.27	302	30.01	84	85	521.100	524.100	Stack 2	0.996
														1.553			681	217.5									
070 AAO HCl 03	1512	1525	3-May	225	±	0.1	22	5	700	0.8	9	84	6.3	1.678	25	582	689	224.7	84.19	302	30.01	84	86	524.400	527.400	Stack 2	0.996
														1.679			689	224.8									
				227	±	10.7							8.4														
													0.4														

Table A-3, continued (Reference Table 3-3)

					Concentration			Reactor Conditions							Analytical					Dry Gas Meter								
WSITE	NSMOOL	Time, Begin	Time, End	Date	Average	±	σ ppm (V) wet @ 3% O2	Condition Number	Flow kscfm	Temp deg F	NH3/NOx	H2O %	Unit 5 Load MW	Sample O2 %	CT ppm	DI	Vol ml	CT µM	HCl ppm wet @ 3% O2	Volume, Total STD L, wet	Temp, Avg. deg K	Barometric Pressure in. Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin ft3	Volume, End ft3	Meter I.D	Corr. Factor
070 BAI HCl 01		1347	1400	4-May	226	±	3.1	22	5	700	0.8	9	84	3.4	1.817	25	677	867	228.6	86.96	296	30.01	74	74	268.500	271.500	Stack 1	1.0080
															1.782			851	224.2									
070 BAI HCl 02		1411	1424	4-May	215	±	0.2	22	5	700	0.8	9	84	3.4	1.872	25	619	817	215.3	86.88	297	30.01	74	75	272.000	275.000	Stack 1	1.0090
															1.870			816	215.3									
070 BAI HCl 03		1435	1449	4-May	223	±	0.7	22	5	700	0.8	9	84	3.7	1.921	25	615	833	223.8	86.90	297	30.01	75	75	275.500	278.500	Stack 1	1.0080
															1.913			830	222.8									
					222	±	5.7							3.5														
														0.2														
070 BAO HCl 01		1347	1404	4-May	225	±	1.0	22	5	700	0.8	9	84	7.7	1.357	25	673	635	226.0	85.29	299	30.01	77	79	538.000	541.000	Stack 2	0.9960
															1.329			631	224.7									
070 BAO HCl 02		1410	1427	4-May	225	±	1.8	22	5	700	0.8	9	84	7.4	1.357	25	672	643	224.0	85.29	299	30.01	78	78	541.400	544.400	Stack 2	0.9960
															1.372			650	226.4									
070 BAO HCl 03		1434	1447	4-May	213	±	4.3	22	5	700	0.8	9	84	7.8	1.347	25	616	585	209.8	83.36	298	30.01	77	78	544.800	547.800	Stack 2	0.9960
															1.386			602	215.9									
					221	±	7.2							7.8														
														0.2														

Table A-3, continued (Reference Table 3-3)

				Time, Begin	Time, End	Date	Concentration ppm (%) wet @ 3% O₂ Average	±	σ	Reactor Conditions Condition Number	Flow kscfm	Temp deg F	NH₃/NO₂	H₂O %	Unit 5 Load MW	Sample O₂ %	Analytical CT ppm	DI	Vol ml	CT µM	HCl ppm wet @ 3% O₂	Dry Gas Meter Volume, Total STD L, wet	Temp, Avg deg K	Barometric Pressure In. Hg	Temp, Begin deg F	Temp, End deg F	Volume, Begin m3	Volume, End m3	Meter I.D.	Corr. Factor
070	CAI	HCl	01	1355	1408	5-May	234	±	0.8	22	5	700	0.8	9	84	4.7	1,777 / 1,783	25	653	818 / 821	233.4 / 234.2	86.77	298	30.14	77	78	292.000	295.000	Stack 1	1.0080
070	CAI	HCl	02	1419	1432	5-May	235	±	0.1	22	5	700	0.8	9	84	4.7	1,909 / 1,910	25	610	821 / 822	234.9 / 235.0	86.53	299	30.14	79	79	295.500	298.500	Stack 1	1.0080
070	CAI	HCl	03	1443	1457	5-May	230	±	0.8	22	5	700	0.8	9	84	3.8	2,014 / 2,024	25	596	847 / 851	229.1 / 230.2	86.65	300	30.14	79	81	299.000	302.010	Stack 1	1.0080
							233	±	2.8							4.4 / 0.5														
070	CAO	HCl	01	1354	1410	5-May	225	±	0.3	22	5	700	0.8	9	84	5.3	1,499 / 1,502	25	703	743 / 745	224.9 / 225.3	84.94	301	30.14	82	83	558.300	561.300	Stack 2	0.9960
070	CAO	HCl	02	1418	1435	5-May	211	±	0.2	22	5	700	0.8	9	84	4.8	1,683 / 1,681	25	605	718 / 717	211.1 / 210.9	84.71	302	30.14	83	85	561.700	564.700	Stack 2	0.9960
070	CAO	HCl	03	1443	1500	5-May	226	±	0.9	22	5	700	0.8	9	84	5.5	1,695 / 1,705	25	612	732 / 736	225.5 / 226.8	84.48	303	30.14	85	86	565.100	568.100	Stack 2	0.9960
							221	±	8.5							5.2 / 0.4														

Table A-4 (Reference Table 3-4)

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAI-M17-01

Flue gas O_2	5.5 % wet	Nozzle diameter	0.469 in.
Flue gas CO_2	13 % wet	Gas meter corr.	1.055
Flue gas N_2	73.66 % wet	Pitot corr. (C_p)	0.821
Flue gas H_2O	7.84 % wet	Duct Area	15.47 ft^2
Ambient Pressure	30.01 in. Hg		
Stack Pressure	-11 in. H_2O		
Stack Temp	683.78 °F		
Volume water	32.1 ml	Meter volume	17.434 ft^3
Particle mass	4437.8 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in. H_2O
Meter Temp	88.94 °F		

FINAL CALCULATED DATA

% WATER	7.84%	gr/ACF :	1.5996 wet
% ISOKINETIC	91.08%	gr/SCF :	3.9134 dry
STACK VEL.	14.49 ft/s	mg/ACM :	3660.59 wet
		mg/SCM :	8955.45 dry
STACK FLOW	13450 ACFM (wet)		
STACK FLOW	5482 SCFM (dry)	LB/MBTU	7.3157

A-13

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAI-M17-02

Flue gas O2	5.5 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.055
Flue gas N2	71.68 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.82 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.01 in Hg		
Stack Pressure	-11 in H2O		
Stack Temp	683.78 °F		
Volume water	41.4 ml	Meter volume	17.61 ft3
Particle mass	4122.8 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in H2O
Meter Temp	89.889 °F		

FINAL CALCULATED DATA

% WATER	9.82%	gr/ACF :	1.4422 wet
% ISOKINETIC	93.54%	gr/SCF :	3.6055 dry
STACK VEL.	14.54 ft/s	mg/ACM :	3300.37 wet
		mg/SCM :	8250.88 dry
STACK FLOW	13495 ACFM (wet)		
STACK FLOW	5382 SCFM (dry)	LB/MBTU	6.7401

A-14

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAI-M17-03

Flue gas O2	5.5 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.055
Flue gas N2	72.28 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.22 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.01 in Hg		
Stack Pressure	-11 in H2O		
Stack Temp	696 °F		
Volume water	39 ml	Meter volume	17.927 ft3
Particle mass	3505.1 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in H2O
Meter Temp	94.7 °F		

FINAL CALCULATED DATA

% WATER	9.22%	gr/ACF :	1.2101 wet
% ISOKINETIC	94.37%	gr/SCF :	3.0375 dry
STACK VEL.	14.60 ft/s	mg/ACM :	2769.09 wet
		mg/SCM :	6950.98 dry
STACK FLOW	13554 ACFM (wet)		
STACK FLOW	5383 SCFM (dry)	LB/MBTU	5.6783

A-15

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAO-M17-01

Flue gas O2	5.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.002
Flue gas N2	73.55 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	7.95 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in Hg		
Stack Pressure	-22.5 in H2O		
Stack Temp	318.44 °F		
Volume water	40.4 ml	Meter volume	22.585 ft3
Particle mass	4856.6 mg	SQRT DP pitot	0.7984
Run time	45 min	AVG DH orifice	0.9522 in H2O
Meter Temp	85.11 °F		

FINAL CALCULATED DATA

% WATER	7.95%	gr/ACF :	2.0117 wet
% ISOKINETIC	97.17%	gr/SCF :	3.4530 dry
STACK VEL.	54.15 ft/s	mg/ACM :	4603.66 wet
		mg/SCM :	7901.93 dry
STACK FLOW	10916 ACFM (wet)		
STACK FLOW	6341 SCFM (dry)	LB/MBTU	6.4551

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAO-M17-02

Flue gas O2	5.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.002
Flue gas N2	72.19 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.31 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in Hg		
Stack Pressure	-22.5 in H2O		
Stack Temp	317.2 °F		
Volume water	52.4 ml	Meter volume	24.92 ft3
Particle mass	5096.7 mg	SQRT DP pitot	0.8626
Run time	45 min	AVG DH orifice	1.11 in H2O
Meter Temp	91.28 °F		

FINAL CALCULATED DATA

% WATER	9.31%	gr/ACF :	1.9088 wet
% ISOKINETIC	99.29%	gr/SCF :	3.3201 dry
STACK VEL.	58.59 ft/s	mg/ACM :	4368.12 wet
		mg/SCM :	7597.76 dry
STACK FLOW	11812 ACFM (wet)		
STACK FLOW	6771 SCFM (dry)	LB/MBTU	6.2066

INPUT DATA

DATE: 4/25/94
RUN ID: 69-AAO-M17-03

Flue gas O2	5.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.002
Flue gas N2	72.56 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	8.94 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.01 in Hg		
Stack Pressure	-22.5 in H2O		
Stack Temp	316.44 °F		
Volume water	50.8 ml	Meter volume	25.36 ft3
Particle mass	4427.7 mg	SQRT DP pitot	0.8577
Run time	45 min	AVG DH orifice	1.096 in H2O
Meter Temp	93.83 °F		

FINAL CALCULATED DATA

% WATER	8.94%	gr/ACF :	1.6453 wet
% ISOKINETIC	100.76%	gr/SCF :	2.8475 dry
STACK VEL.	58.19 ft/s	mg/ACM :	3765.00 wet
		mg/SCM :	6516.19 dry
STACK FLOW	11732 ACFM (wet)		
STACK FLOW	6758 SCFM (dry)	LB/MBTU	5.3231

INPUT DATA

DATE: 5/2/94
RUN ID: 70-BAI-M17-01

Flue gas O2	3.5 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.055
Flue gas N2	72.2 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.3 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-14 in H2O		
Stack Temp	689.89 °F		
Volume water	33.9 ml	Meter volume	14.683 ft3
Particle mass	3422 mg	SQRT DP pitot	0.1549
Run time	45 min	AVG DH orifice	0.38 in H2O
Meter Temp	68.89 °F		

FINAL CALCULATED DATA

% WATER	9.30%	gr/ACF :	1.3716 wet
% ISOKINETIC	94.18%	gr/SCF :	3.4431 dry
STACK VEL.	12.61 ft/s	mg/ACM :	3138.76 wet
		mg/SCM :	7879.23 dry
STACK FLOW	11703 ACFM (wet)		
STACK FLOW	4634 SCFM (dry)	LB/MBTU	5.6967

INPUT DATA

DATE:	**5/2/94**		
RUN ID:	**70-BAI-M17-02**		

Flue gas O2	3.5 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.055
Flue gas N2	71.83 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.67 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-14 in H2O		
Stack Temp	687.33 °F		

Volume water	35.1 ml	Meter volume	14.846 ft3
Particle mass	3240.8 mg	SQRT DP pitot	0.1549
Run time	45 min	AVG DH orifice	0.38 in H2O
Meter Temp	79.17 °F		

FINAL CALCULATED DATA

% WATER	9.67%	gr/ACF :	1.3073 wet
% ISOKINETIC	93.62%	gr/SCF :	3.2877 dry
STACK.VEL.	12.60 ft/s	mg/ACM :	2991.57 wet
		mg/SCM :	7523.62 dry
STACK FLOW	11697 ACFM (wet)		
STACK FLOW	4623 SCFM (dry)	LB/MBTU	5.4396

INPUT DATA

DATE:	5/2/94		
RUN ID:	70-BAI-M17-03		

Flue gas O2	3.5 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.055
Flue gas N2	70.35 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	11.15 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-14 in H2O		
Stack Temp	689.11 °F		
Volume water	41.1 ml	Meter volume	14.983 ft3
Particle mass	3479.2 mg	SQRT DP pitot	0.1549
Run time	45 min	AVG DH orifice	0.38 in H2O
Meter Temp	84.83 °F		

FINAL CALCULATED DATA

% WATER	11.15%	gr/ACF :	1.3800 wet
% ISOKINETIC	94.90%	gr/SCF :	3.5340 dry
STACK VEL.	12.64 ft/s	mg/ACM :	3158.06 wet
		mg/SCM :	8087.28 dry
STACK FLOW	11735 ACFM (wet)		
STACK FLOW	4555 SCFM (dry)	LB/MBTU	5.8471

Table A-4, continued (Reference Table 3-4)

INPUT DATA

DATE: 5/2/94
RUN ID: 70-BAO-M17-01

Flue gas O2	3.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.002
Flue gas N2	72.59 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	8.91 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-20 in H2O		
Stack Temp	317.33 °F		
Volume water	46.7 ml	Meter volume	22.585 ft3
Particle mass	3837.4 mg	SQRT DP pitot	0.7703
Run time	45 min	AVG DH orifice	0.904 in H2O
Meter Temp	75.61 °F		

FINAL CALCULATED DATA

% WATER	8.91%	gr/ACF :	1.5582 wet
% ISOKINETIC	103.41%	gr/SCF :	2.6731 dry
STACK VEL.	51.91 ft/s	mg/ACM :	3565.87 wet
		mg/SCM :	6117.18 dry
STACK FLOW	10465 ACFM (wet)		
STACK FLOW	6064 SCFM (dry)	LB/MBTU	4.4227

A-22

INPUT DATA

DATE: 5/2/94
RUN ID: 70-BAO-M17-02

Flue gas O2	3.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.002
Flue gas N2	72.14 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.36 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-20 in H2O		
Stack Temp	315.78 °F		
Volume water	50.1 ml	Meter volume	23.343 ft3
Particle mass	3911.2 mg	SQRT DP pitot	0.7898
Run time	45 min	AVG DH orifice	0.9489 in H2O
Meter Temp	85.11 °F		

FINAL CALCULATED DATA

% WATER	9.36%	gr/ACF :	1.5590 wet
% ISOKINETIC	102.76%	gr/SCF :	2.6826 dry
STACK VEL.	53.21 ft/s	mg/ACM :	3567.71 wet
		mg/SCM :	6138.76 dry
STACK FLOW	10728 ACFM (wet)		
STACK FLOW	6198 SCFM (dry)	LB/MBTU	4.4383

Table A-4, continued (Reference Table 3-4)

INPUT DATA

DATE: 5/2/94
RUN ID: 70-BAO-M17-03

Flue gas O2	3.5 % wet	Nozzle diameter	0.219 inches
Flue gas CO2	15 % wet	Gas meter corr.	1.002
Flue gas N2	71.73 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	9.77 % wet	Duct Area	3.36 ft2
Ambient Pressure	30.1 in Hg		
Stack Pressure	-20 in H2O		
Stack Temp	315.22 °F		
Volume water	52.4 ml	Meter volume	23.527 ft3
Particle mass	4242.7 mg	SQRT DP pitot	0.7948
Run time	45 min	AVG DH orifice	0.962 in H2O
Meter Temp	90.78 °F		

FINAL CALCULATED DATA

% WATER	9.77%	gr/ACF :	1.6889 wet
% ISOKINETIC	102.21%	gr/SCF :	2.9171 dry
STACK VEL.	53.57 ft/s	mg/ACM :	3864.93 wet
		mg/SCM :	6675.53 dry
STACK FLOW	10799 ACFM (wet)		
STACK FLOW	6215 SCFM (dry)	LB/MBTU	4.8264

INPUT DATA

DATE: 4/26/94
RUN ID: 69-CAI-M17-01

Flue gas O2	4 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.055
Flue gas N2	71.88 % wet	Pitot corr. (Cp)	0.821
Flue gas.H2O	11.12 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.02 in Hg		
Stack Pressure	-11 in H2O		
Stack Temp	686.1 °F		
Volume water	47.5 ml	Meter volume	17.597 ft3
Particle mass	4154.9 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in H2O
Meter Temp	90.67 °F		

FINAL CALCULATED DATA

% WATER	11.12%	gr/ACF :	1.4326 wet
% ISOKINETIC	94.49%	gr/SCF :	3.6402 dry
STACK VEL.	14.60 ft/s	mg/ACM :	3278.45 wet
		mg/SCM :	8330.32 dry
STACK FLOW	13553 ACFM (wet)		
STACK FLOW	5316 SCFM (dry)	LB/MBTU	6.2010

INPUT DATA

DATE: 4/26/94
RUN ID: 69-CAI-M17-02

Flue gas O2	4 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.055
Flue gas N2	72.41 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	10.59 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.02 in Hg		
Stack Pressure	-11 in H2O		
Stack Temp	682 °F		
Volume water	45.4 ml	Meter volume	17.843 ft3
Particle mass	4480.1 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in H2O
Meter Temp	93.11 °F		

FINAL CALCULATED DATA

% WATER	10.59%	gr/ACF :	1.5449 wet
% ISOKINETIC	94.74%	gr/SCF :	3.8882 dry
STACK VEL.	14.56 ft/s	mg/ACM :	3535.26 wet
		mg/SCM :	8897.76 dry
STACK FLOW	13516 ACFM (wet)		
STACK FLOW	5353 SCFM (dry)	LB/MBTU	6.6234

INPUT DATA

DATE: 4/26/94
RUN ID: 69-CAI-M17-03

Flue gas O2	4 % wet	Nozzle diameter	0.469 inches
Flue gas CO2	13 % wet	Gas meter corr.	1.055
Flue gas N2	70.38 % wet	Pitot corr. (Cp)	0.821
Flue gas H2O	12.62 % wet	Duct Area	15.47 ft2
Ambient Pressure	30.02 in Hg		
Stack Pressure	-11 in H2O		
Stack Temp	683.67 °F		
Volume water	54.8 ml	Meter volume	17.869 ft3
Particle mass	4537.9 mg	SQRT DP pitot	0.1789
Run time	45 min	AVG DH orifice	0.54 in H2O
Meter Temp	99.44 °F		

FINAL CALCULATED DATA

% WATER	12.62%	gr/ACF :	1.5423 wet
% ISOKINETIC	95.72%	gr/SCF :	3.9777 dry
STACK VEL.	14.62 ft/s	mg/ACM :	3529.43 wet
		mg/SCM :	9102.50 dry
STACK FLOW	13573 ACFM (wet)		
STACK FLOW	5246 SCFM (dry)	LB/MBTU	6.7758

A-27
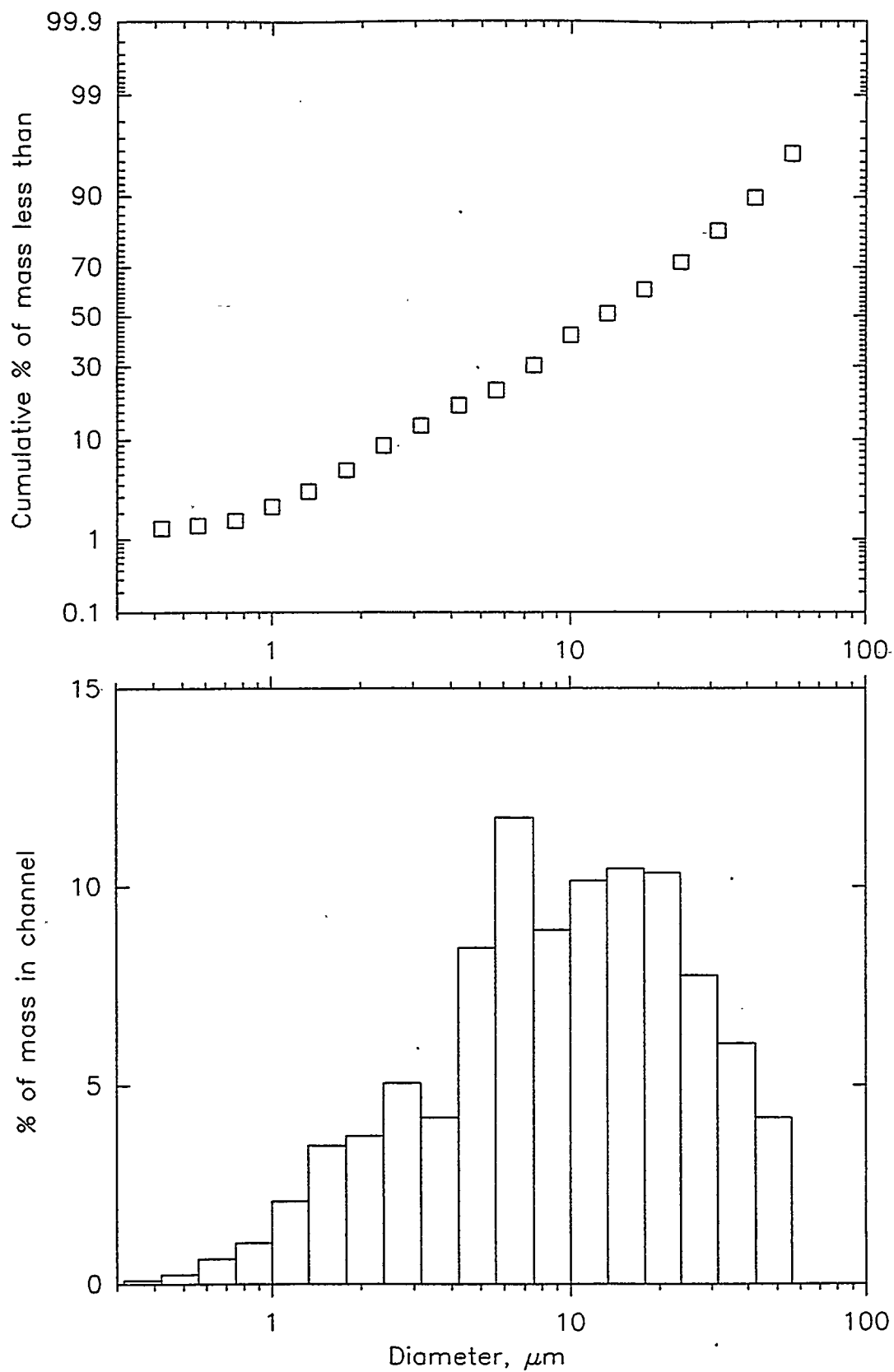
Figure A-1. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor A air heater inlet.
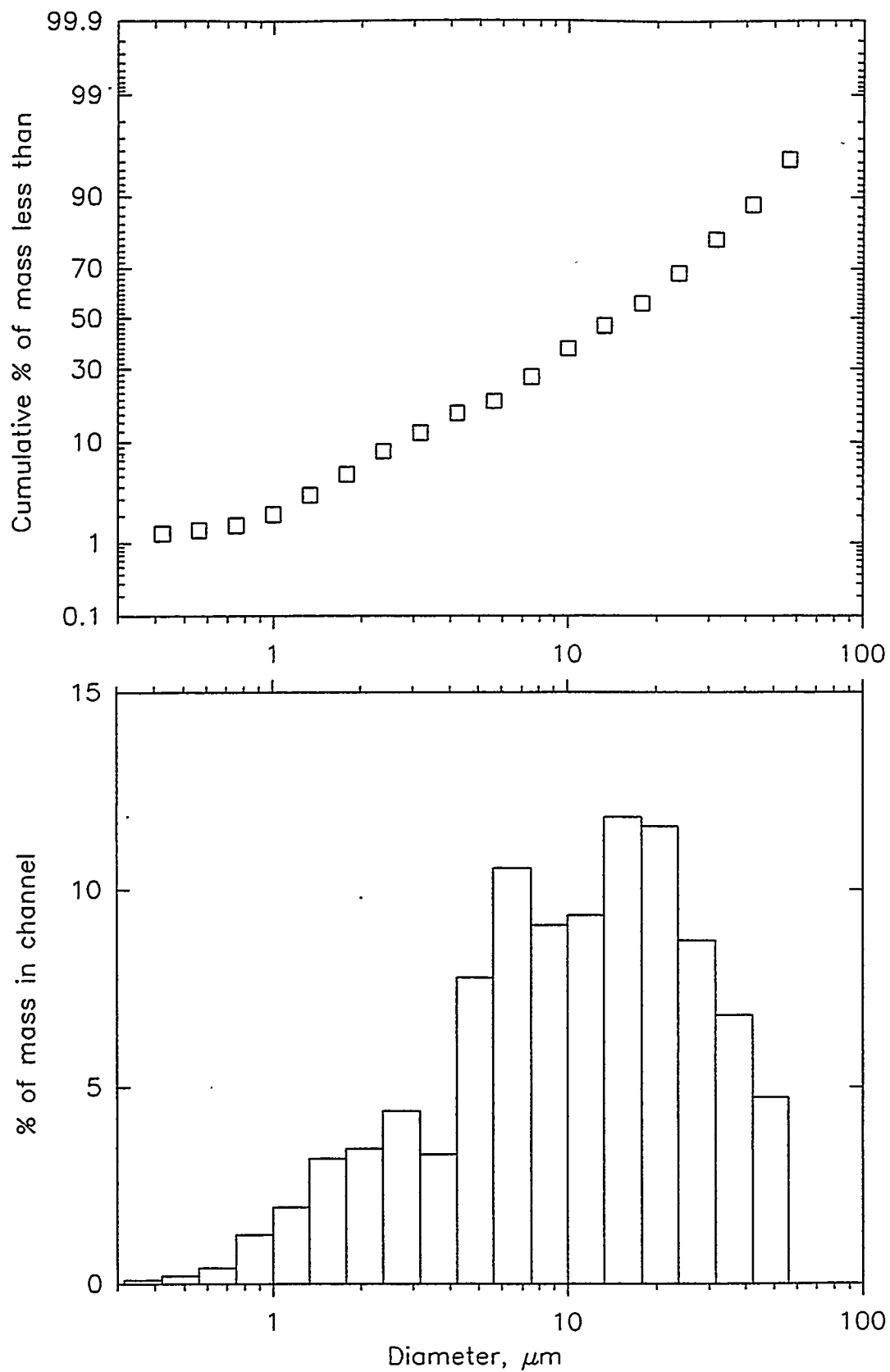
Figure A-2. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor A air heater outlet.
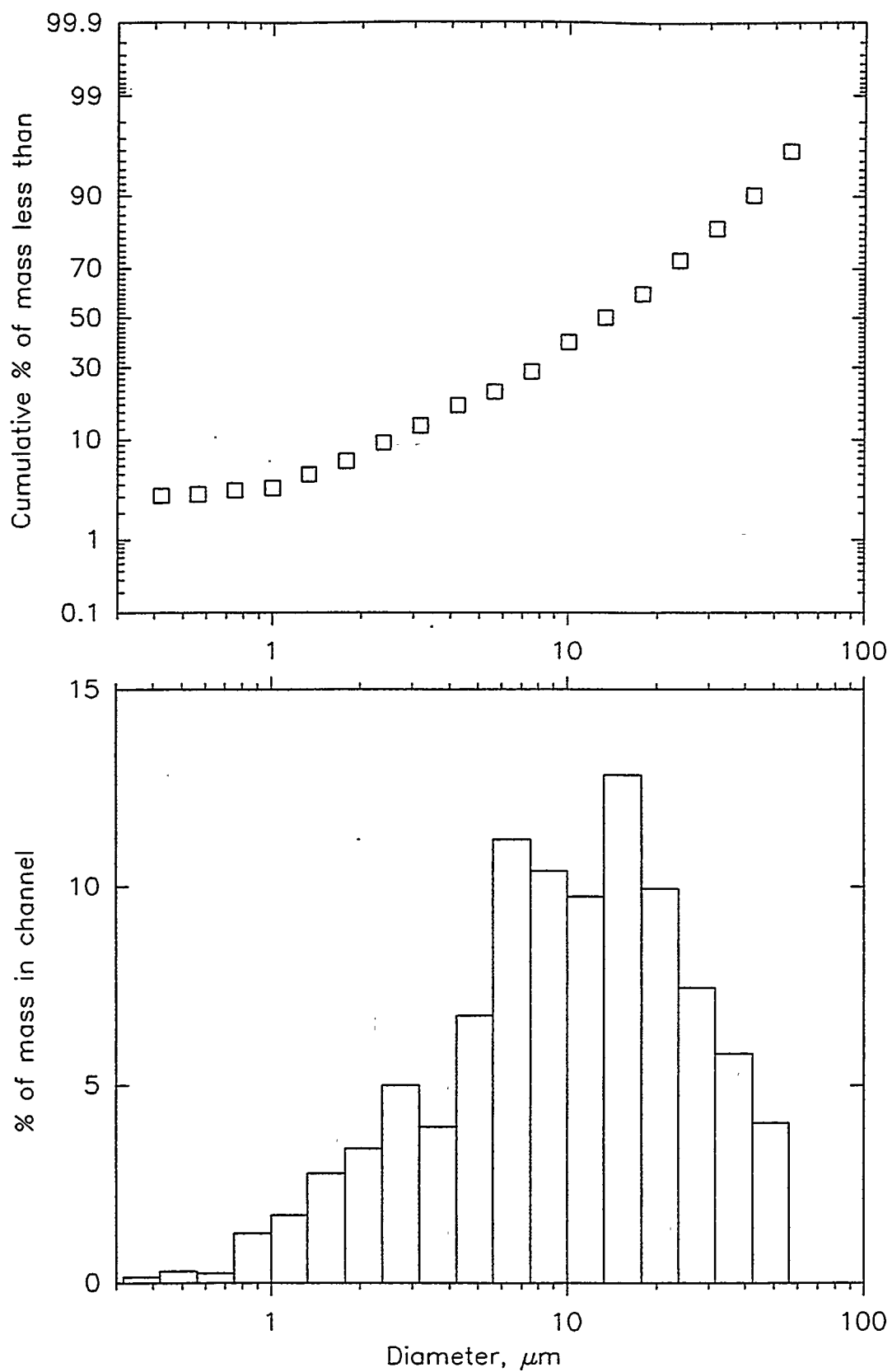
Figure A-3. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor B air heater inlet.
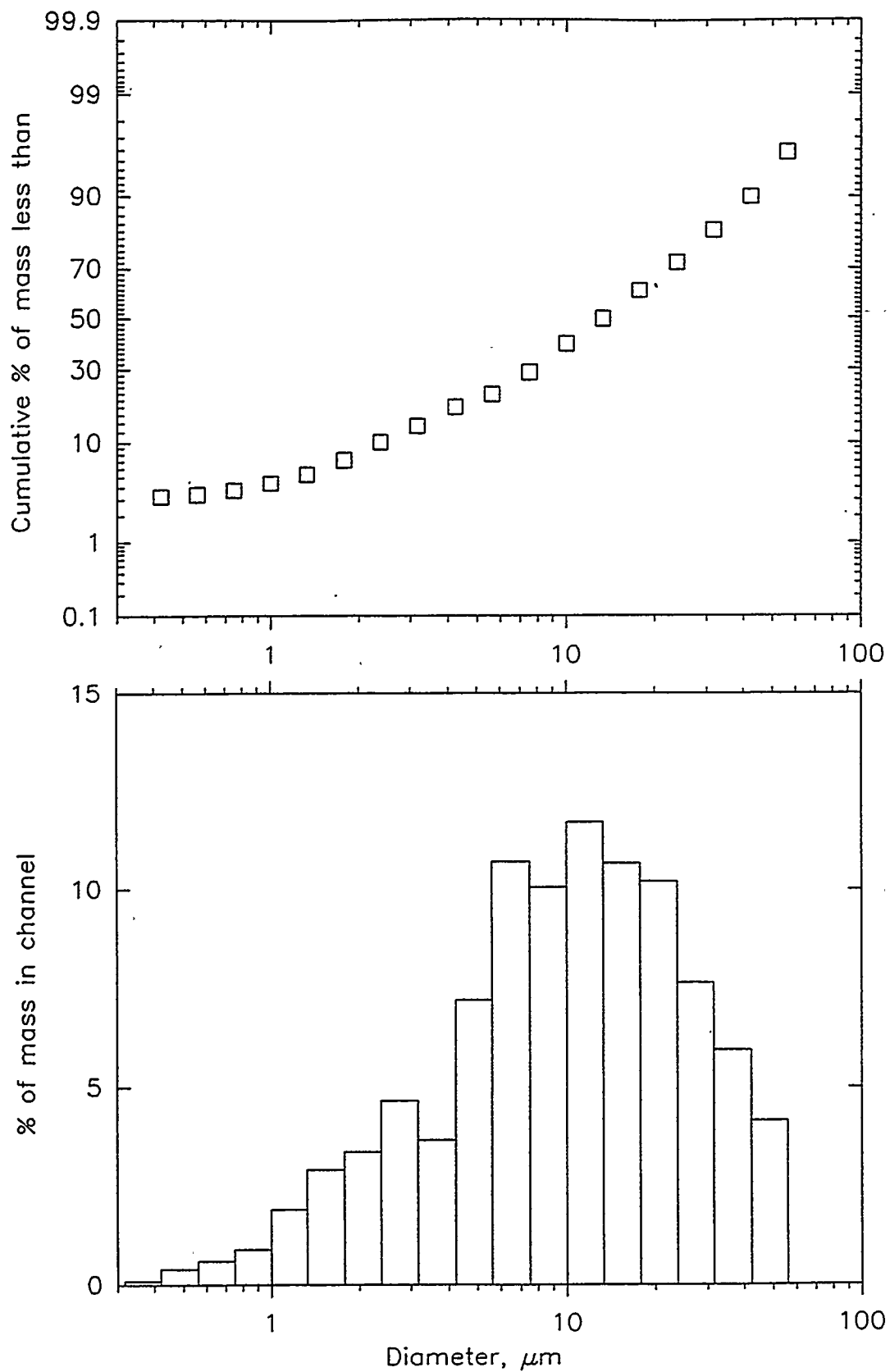
Figure A-4. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor B air heater outlet.
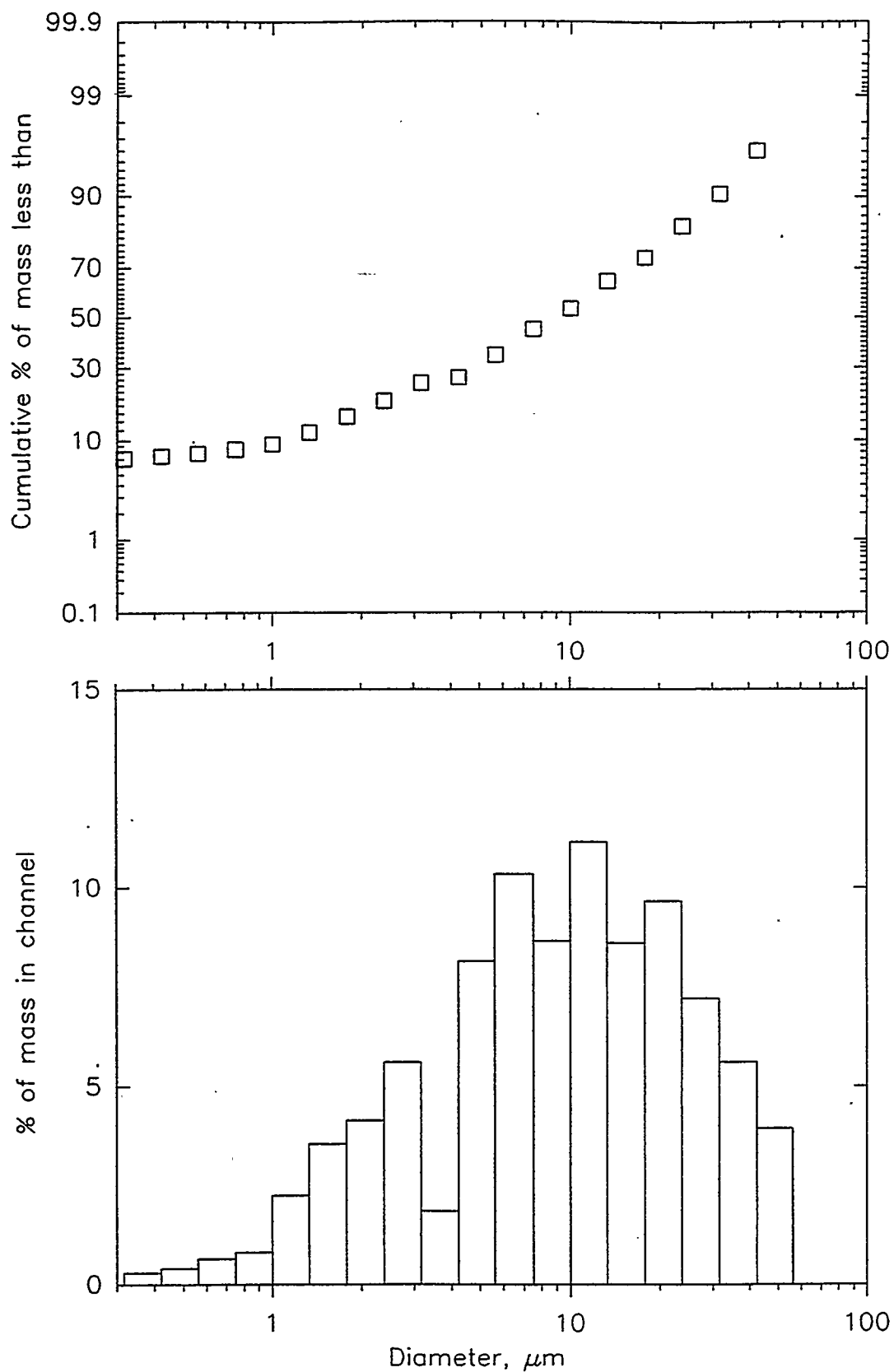
Figure A-5. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor C air heater inlet.

RESISTIVITY TEST DATA

START DATE _23 SEPT 94_ APPARATUS NO. ___2___   TEST TYPE:
PROJECT NO. _7613.07.01_   DESCENDING _✓455°C_
TEST NO. _____   ASCENDING _____
PLANT/COMPANY ___PLANT CRIST UNIT 5___   ACID _____

Position	Cell No.	Cell Volume	Weight Full	Weight Empty	Weight Ash	Ash Density	Ash Volume	%P
1	1							
2	2							
3	3							
4	4							

Position	Ash I.D.	Thickness	Sink	A/(T-S)	Mesh Size
1	060-E4-M17-01,02			8.00	-60
2	060-F4-M17-01,02	.495cm	.2cm	13.15	"
3	046-G4-M17-01,2,3			7.92	"
4	No. 4090 A INLET			7.84	"

Thermocouple Readings in °C

Time ___ ___ ___   ___ ___ ___   ___ ___ ___
 1 ___ ___ ___   5 ___ ___ ___   9 ___ ___ ___
 2 ___ ___ ___   6 ___ ___ ___   10 ___ ___ ___
 3 ___ ___ ___   7 ___ ___ ___   11 ___ ___ ___
 4 ___ ___ ___   8 ___ ___ ___   12 ___ ___ ___

Gas Composition: _____

Water Bath Temperature: _44.5_

Acid Bath Temperature: _____

Volume % Water:

Weight Wet: _____

Weight Dry: _____

Weight Water: _____

$$\frac{(\quad)(1.244) \times 100}{(\quad)(1.244) + [\quad \cdot 0.8 \cdot (460/\quad)]} = 8.7\%$$

FORM ENV021   SOUTHERN RESEARCH INSTITUTE

RESISTIVITY TEST DATA

START DATE **23 SEPT 94**   APPARATUS NO. **2**   PAGE NO. **1**

PROJECT NO. **7613.07.01**

TEST SAMPLE I.D. **PLANT CRIST UNIT 5**

Date/Time	Temperature °C	Temperature 1000/°K	D.C. Volts	D.C. Amperes	A/L	Resistivity Ohm-cm	Comments
23SEPT/0833	440/442	1.40	-2000	.10 x10^-3	8.00	1.60 x10^8	0750-H2O ON
	434/438	1.41	-200 ★	.10 x10^-4	13.15	2.63 x10^8	0821-OVEN OFF
	431/436	.11	-2000	.10 x10^-3	7.92	1.58 x10^8	★P.S TRIPS @ -2KV
	428/434	"	- "	.18 x10^-3	7.84	7.71 x10^7	
0856	384/398	1.49	-2000	.28 x10^-4		5.71 x10^8	
	380/394	1.50	-200	.36 x10^-5		7.31 x10^8	
	377/391	1.51	-2000	.29 x10^-4		5.46 x10^8	
	374/388	"	"	.53 x10^-4		2.96 x10^8	
0914	340/357	1.59	-2000	.22 x10^-5		1.?? x10^9	
	338/355	"	-200	.15 x10^-5		1.75 x10^9	
	330/347	1.61	-2000	.76 x10^-5		2.07 x10^9	
	327/344	1.62	"	.135 x10^-4		1.16 x10^9	
0942	283/301	1.74	-2000	.14 x10^-5		1.14 x10^10	
	280/298	1.75	-200	.34 x10^-6		7.74 x10^9	
	277/295	1.76	-2000	.13 x10^-5		1.22 x10^10	
	275/293	1.77	"	.24 x10^-5		6.53 x10^9	
1006	243/261	1.87	-2000	.35 x10^-6		4.57 x10^10	
	241/259	1.88	-200	.11 x10^-6		2.39 x10^10	
	239/257	1.89	-2000	.33 x10^-6		4.80 x10^10	
	237/255	"	"	.57 x10^-6		2.75 x10^10	
1025	216/233	1.98	-2000	.14 x10^-6		1.14 x10^11	
	214/231	"	-200	~.68 x10^-7 ★★		3.87 x10^10	★★ SLIGHTLY ERRATIC
	212/229	1.99	-2000	.13 x10^-6		1.22 x10^11	
	211/228	2.00	"	.21 x10^-6		7.47 x10^10	
1104	172/188	2.17	-2000	.38 x10^-7		4.21 x10^11	
	171/187	"	-200	.19 x10^-7		1.38 x10^11	
	169/185	2.18	-2000	.39 x10^-7		4.06 x10^11	
	168/184	2.19	"	.48 x10^-7		3.27 x10^11	
				x10^-		x10	
				x10^-		x10	

RESISTIVITY TEST DATA

START DATE _23 SEPT 94_ APPARATUS NO. _2_ PAGE NO. _2_

PROJECT NO. _7613.17.01_

TEST SAMPLE I.D. _PLANT CRIST UNIT 5_

Date/Time	Temperature °C	1000/°K	D.C. Volts	D.C. Amperes	A/L	Resistivity Ohm-cm	Comments
1140	142/156	2.33	−2000	$.24 \times 10^{-7}$	8.00	6.67×10^{11}	
	141/155	2.34	−200	$.24 \times 10^{-7}$	13.15	1.10×10^{11}	
	140/154	"	−2000	$.28 \times 10^{-7}$	7.92	5.66×10^{11}	
	139/153	2.35	"	$.26 \times 10^{-7}$	7.84	6.03×10^{11}	
1238	108/119	2.55	−2000	$.49 \times 10^{-7}$		3.27×10^{11}	
	106/117	256	−200	$.87 \times 10^{-7}$		3.02×10^{10}	
	107/118	"	−2000	$.66 \times 10^{-7}$		2.40×10^{11}	
	106/117	"	"	$.50 \times 10^{-7}$		3.14×10^{11}	
1353	81/90	2.75	−2000	$.63 \times 10^{-6}$		2.54×10^{10}	
	79/89	2.76	−200	$.28 \times 10^{-6}$		9.39×10^{9}	HIGHEST CARBON CONTENT AFTER
	80/89	"	−2000	$.97 \times 10^{-6}$		1.63×10^{10}	TEST. ALSO EVIDENCE OF WINDER COATING
	80/89	"	"	$.67 \times 10^{-6}$		2.34×10^{10}	THINNER LAYER
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	

RESISTIVITY TEST DATA

START DATE _27 SEPT 94_ APPARATUS NO. ___2___ TEST TYPE:
PROJECT NO. _7613.07.01_ DESCENDING _✓ 455°C_
TEST NO._____ ASCENDING_____
PLANT/COMPANY _____PLANT CRIST UNIT 5_____ ACID_____

Position	Cell No.	Cell Volume	Weight Full	Weight Empty	Weight Ash	Ash Density	Ash Volume	%P
1.	1							
2	2							
3	3							
4	4							

Position	Ash I.D.	Thickness	Sink	A/(T−S)	Mesh Size
1	No. 4091 A OUTLET			8.00	−60
2	4092 B INLET			7.84	''
3	4093 B OUTLET			7.92	''
4	4094 C INLET			7.84	''

Thermocouple Readings in °C

Time ____ ____ ____   ____ ____ ____   ____ ____ ____
1 ____ ____ ____   5 ____ ____ ____   9 ____ ____ ____
2 ____ ____ ____   6 ____ ____ ____   10 ____ ____ ____
3 ____ ____ ____   7 ____ ____ ____   11 ____ ____ ____
4 ____ ____ ____   8 ____ ____ ____   12 ____ ____ ____

Gas Composition: _____

Water Bath Temperature: _____

Acid Bath Temperature: _____

Volume % Water:

Weight Wet: _____

Weight Dry: _____

Weight Water: _____

$$\frac{(\quad)(1.244) \times 100}{(\quad)(1.244) + [\quad \cdot 0.8 \cdot (460/\quad)]} = \quad \%$$

RESISTIVITY TEST DATA

START DATE _27 SEPT 94_ APPARATUS NO. __2__ PAGE NO. ___1___

PROJECT NO. _7613.07.0_

TEST SAMPLE I.D. ___PLANT CRIST UNIT 5___

Date/Time	Temperature °C	Temperature 1000/°K	D.C. Volts	D.C. Amperes	A/L	Resistivity Ohm-cm	Comments
27SEPT/0831	443/444	1.39	−200	$.12 \times 10^{-4}$	8.00	$\times 10$	0750 – H₂O 0~
	441/442	1.40	"	$.21 \times 10^{-4}$	7.74	$\times 10$	
	439/440	"	"	$.19 \times 10^{-4}$	7.92	$\times 10$	
	436/437	1.41	"	$.14 \times 10^{-4}$	7.84	$\times 10$	
0835	433/433	1.42	−2000	$.15 \times 10^{-3}$		1.07×10^{8}	
	431/432	"	"	$.28 \times 10^{-3}$		5.60×10^{7}	
	427/428	1.43	"	$.23 \times 10^{-3}$		6.89×10^{7}	
	425/426	"	"	$.18 \times 10^{-3}$		8.71×10^{7}	
0858	377/323	1.52	"	$.42 \times 10^{-4}$		3.81×10^{8}	
	377/375	1.53	"	$.70 \times 10^{-4}$		2.24×10^{8}	
	371/377	1.54	"	$.58 \times 10^{-4}$		2.73×10^{8}	
	367/374	1.55	"	$.43 \times 10^{-4}$		3.65×10^{8}	
0925	315/332	1.65	"	$.74 \times 10^{-5}$		2.16×10^{9}	
	313/330	1.66	"	$.115 \times 10^{-4}$		1.36×10^{9}	
	311/328	"	"	$.10 \times 10^{-4}$		1.58×10^{9}	
	308/325	1.67	"	$.73 \times 10^{-5}$		2.15×10^{9}	
0944	278/296	1.76	"	$.21 \times 10^{-5}$		7.62×10^{9}	
	276/294	"	"	$.33 \times 10^{-5}$		4.75×10^{9}	
	274/292	1.77	"	$.29 \times 10^{-5}$		5.46×10^{9}	
	272/290	1.78	"	$.205 \times 10^{-5}$		7.65×10^{9}	
1003	246/264	1.86	"	$.63 \times 10^{-6}$		2.54×10^{10}	
	244/262	1.87	"	$.10 \times 10^{-5}$		1.57×10^{10}	
	242/260	1.88	"	$.89 \times 10^{-6}$		1.78×10^{10}	
	240/258	"	"	$.63 \times 10^{-6}$		2.49×10^{10}	
1023	217/234	1.97	"	$.21 \times 10^{-6}$		7.62×10^{10}	
	215/232	1.98	"	$.34 \times 10^{-6}$		4.61×10^{10}	
	215/232	"	"	$.30 \times 10^{-6}$		5.28×10^{10}	
	213/230	1.99	"	$.22 \times 10^{-6}$		7.13×10^{10}	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	

RESISTIVITY TEST DATA

START DATE _27 SEPT 94_   APPARATUS NO. _2_   PAGE NO. _2_
PROJECT NO. _7613.07.01_
TEST SAMPLE I.D. _PLANT CRIST UNIT 5_

Date/Time	Temperature °C	Temperature 1000/°K	D.C. Volts	D.C. Amperes	A/L	Resistivity Ohm-cm	Comments
1102	173/189	2.16	−2000	$.39 \times 10^{-7}$	8.00	4.10×10^{11}	
	172/188	2.17	!	$.70 \times 10^{-7}$	7.84	2.24×10^{11}	
	171/186	2.18	:	$.63 \times 10^{-7}$	7.92	2.51×10^{11}	
	170/185	''		$.52 \times 10^{-7}$	7.84	3.02×10^{11}	
1142	140/154	2.34		$.155 \times 10^{-7}$		1.03×10^{12}	
	139/153	2.35		$.35 \times 10^{-7}$		4.48×10^{11}	
	138/152	''		$.28 \times 10^{-7}$		5.66×10^{11}	
	138/152	''		$.29 \times 10^{-7}$		5.41×10^{11}	
1236	109/120	2.54		$.20 \times 10^{-7}$		7.01×10^{11}	
	108/119	2.55		$.62 \times 10^{-7}$		2.53×10^{11}	
	107/118	2.56		$.41 \times 10^{-7}$		3.85×10^{11}	
	107/118	''		$.58 \times 10^{-7}$		2.70×10^{11}	
1350	82/87	2.78		$.22 \times 10^{-6}$		2.27×10^{10}	
	81/86	2.79		$.83 \times 10^{-6}$		1.83×10^{10}	
	81/86	''		$.50 \times 10^{-6}$		3.17×10^{10}	
	81/86	''		$.82 \times 10^{-6}$		1.91×10^{10}	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
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				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	
				$\times 10^{-}$		$\times 10$	

Final Test Report on Three Air Preheaters on an SCR DeNO$_x$ Demonstration Project at Plant Crist

Engineering Report 96-2
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I. Introduction

A. Test Background

From May 1993 through July 1995, two Ljungstrom® rotary regenerative air heaters and one Q-Pipe® heat pipe air heater were operated downstream of several SCR DeNOx reactors as part of a DOE Clean Coal Demonstration Project at Southern Company's Plant Crist #5. Performance data were collected during this period to see if there were any significant long term effects on the operating characteristics of the air heaters due to fouling of their heat transfer surfaces. These data are important because the sulfur compounds and the ammonia contained in the gas leaving an SCR reactor can combine to form ammonia bisulfate (ABS), which can condense on the heat transfer surfaces in the air heater to form a solid at normal air heater temperatures. This results in fouling and plugging of the heat transfer surfaces in the air heater. Aggressive cleaning techniques such as water washing and frequent sootblowing of the air heaters were used to clean the heat transfer surfaces.

B. Description of Air Heaters

The two rotary regenerators are identified as air heaters (A/H) A and B. The total depth of element in A/H A is 72 inches, divided into a 30 in. hot end (HE) layer and a 42 in. cold end (CE) layer. A/H B also has a total element depth of 72 inches, but it is divided into three separate layers – an 18 in. HE layer, a 42 in. intermediate layer, and a 12 in. CE layer. The reason for testing both a two and a three layer design is that the three layer arrangement is more traditional because it allows the 12 inch cold end layer to be easily replaced if it suffers excessive corrosion. On the other hand, it was also believed that the discontinuity that exists at the leading edges of the CE layer of A/H B could be more conducive to the accumulation of ABS deposits than the continuous channels of the 42 in. CE layer of A/H A. There was also some question as to which design would be easier to clean with sootblowers, since it was believed that the kinetic energy of the sootblower jet would be severely attenuated by the gap between the CE layer and the intermediate layer of the 3 layer A/H. If the inter-layer gap actually does dissipate the kinetic energy of the sootblower jet, then using a 42 in. CE layer would allow more element to be cleaned by the same sootblowers than would a 12 in. CE layer.

The Q-Pipe® air heater was designated as A/H C, and it contained an array of finned, 2 inch OD in-line heat pipes. On the gas side of each heat pipe, there were three solid fins per inch, the fin OD was 3.5 inches, and the fin thickness was 0.059 in. The centerline spacing between tubes was 3.75 inches in the two directions perpendicular to the tube axis.

C. Effects of Fouling and Cleaning on Regenerators and Recuperators

To knowledgeably compare a regenerative air heater to a recuperative air heater requires some understanding of the fundamental differences between the two, especially insofar as their performance is affected by a certain amount of fouling (i.e., a deposit on the heat transfer surface). The thermal performance of a regenerative air heater tends to be less sensitive to fouling because the heat transfer process in a regenerator mainly just involves convection of heat into and out of the heat transfer surface. There does need to be a certain amount of heat storage by the heat transfer surface, but this can be accomplished by almost any type of mass. It is not

necessary for the heat to be conducted completely through the full thickness of the heat transfer surface, as in a recuperator. Thus, when the heat transfer surface of a regenerator is coated by a layer of deposit, the deposit itself can act as the new heat transfer surface. As long as the gas and air can still flow over the deposit, convection (and heat storage) will still take place, and the regenerator will still function well, albeit not quite at its full design capacity. The thermal conductivity of the deposit has a small impact on the performance of a regenerative air heater. If a regenerator contains a "closed channel"* heat transfer surface that becomes plugged, then the performance of the regenerator should decline noticeably because neither the air nor the gas would be able to reach any of the heat transfer surface inside the plugged channels anywhere along their length. The 42 in. CE layer of A/H A is a closed channel surface, as are the 12 in. CE layer and 42 in. intermediate layer of A/H B.

A recuperator is much less tolerant of deposits on its heat transfer surface than a regenerator because the deposit layer tends to act as an insulator that inhibits the flow of heat into the heat transfer surface. If a layer of deposit covers some part of the heat transfer surface of a recuperator, the heat <u>must</u> be conducted completely through that deposit in order to get to the heat transfer surface. Thus, the thermal conductivity of the deposit layer will have a strong effect on the performance of a recuperator.

The effects of sootblowing are also believed to be different for a Ljungstrom® regenerator compared to a finned tube recuperator. The particular heat transfer surfaces used in the cold end layers of air heaters A and B (and the intermediate layer of A/H B) are more easily cleaned by sootblowing than is an array of finned tubes, because they allow the sootblower jet to pass straight through the flow channels. A finned tube array is difficult to clean with a sootblower because the tube row nearest the sootblower acts as a shield to deflect the sootblower jet before it can reach the other rows in the array.

II. Measurements

Since an air heater's most important performance parameters are its thermal effectiveness and its air and gas side pressure drops, these were the parameters that were emphasized in planning the data acquisition system and in analyzing the data. The amount of air to gas leakage that occurs in a Ljungstrom® air heater is needed to do the performance calculation, but it was felt that the leakage rate itself was not a key component of the analysis, since it was expected to be independent of the fouling that was expected as a result of the ABS condensation.

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* A closed channel heat transfer surface is one wherein flow channels are <u>only</u> open to flow at each end so that any fluid that enters that channel is confined to that channel for the entire length of that layer of heat transfer surface. An open channel heat transfer surface allows the fluid to enter or leave the channel at numerous places along its length.

The primary goal of these A/H tests was to measure the net effect on thermal performance and pressure drop of the ABS fouling after mitigation by the cleaning process. In order to calculate the effectiveness of the air heaters, the following parameters were measured for all three air heaters (except where noted):

1. Air inlet flow rate
2. Gas inlet flow rate
3. Air inlet temperature
4. Gas inlet temperature
5. Air outlet temperature
6. Corrected gas outlet temperature (A and B only)
7. Gas outlet temperature (C)

The corrected gas outlet temperature is defined to be the temperature of the gas leaving the air heater after it has mixed with the air that is allowed to leak into the gas stream by the air heater. Using the corrected gas outlet temperature and the amount of air to gas leakage (A/H's A and B only) allows a calculation of the uncorrected gas outlet temperature, which is needed to determine the capacity ratio for the two regenerative A/H's. Note that since A/H C does not leak, it only has a gas outlet temperature, not a corrected and an uncorrected gas outlet temperature. To know the pressure drops of all three A/H's (except for the air side pressure drop of A/H C) and the leakages of A/H's A and B, the following parameters were measured:

1. Gas inlet O_2 concentrations (A, B, and C)
2. Gas outlet O_2 concentrations (A and B)
3. Gas side pressure drops (A, B, and C)
4. Air side pressure drops (A and B)

The air side pressure drop of A/H C was not of interest because it was never exposed to any gas flow. On the other hand, the air side pressure drops of A/H's A and B were expected to increase over time as a result of fouling and plugging in those A/H's, because the same heat transfer surface is alternately exposed to both the air and gas flows as the rotor rotates from one side to the other. Thus, if the heat transfer surface in either A/H A or B becomes fouled, it would be expected that both the air and the gas side pressure drops for that air heater would rise.

III. Data Analysis

A. General

The purpose of the data analysis is to use the measured data to present a meaningful picture of the steady state performance trends for the three A/H's. The degree to which the performance of an A/H changes over time would then indicate the net effect of both fouling and cleaning on that A/H. Thus, it is necessary that the performance parameters be chosen so that, as much as possible, they will accurately reflect performance changes rather than some other change, such as flow. To accomplish this, the parameters Ntu (Number of transfer units) and Euler number (a measure of flow resistance, i.e., pressure drop divided by velocity head) were selected as measures of thermal performance and pressure drop, respectively, since these two dimensionless parameters are least sensitive to flow rate. As an additional precaution

3

against the inadvertent inclusion of flow effects, it was decided to only include those data points with a gas flow between 4900 and 5100 scfm. Thus, while there may be some impact on Ntu and Eu from the changes in air flow over time, they are believed to be minimal.

B. Data Selection

The large quantity of data that was taken for the three air heaters from May, 1993 through July, 1995 was selectively reduced down to daily averages of the following parameters:

1. Gas inlet flow rate
2. Air inlet flow rate
3. Gas inlet temperature
4. Air inlet temperature
5. Gas outlet temperature
6. Corrected gas outlet temperature (A and B only)
7. Gas outlet temperature (C)
8. Gas inlet O_2 concentration
9. Gas outlet O_2 concentration (A and B only)
10. Gas side pressure drop
11. Air side pressure drop (A and B only)

Only some of the measured data was selected for use in the daily averages, since only the steady state performance of the air heaters was being analyzed. All data taken during or soon after a sootblowing event, a flow change, or a large-temperature change were eliminated because these events produced performance transients. This eliminated temperature and pressure drop spikes or dips that were flow related. In the graphs that accompany this report, each data point represents the average value for a single day. If a single day's average value was obtained from fewer than five steady state readings, that data point was eliminated for statistical reasons.

C. Calculation of Leakage and Uncorrected Gas Out Temperatures for A/H's A & B

As discussed in the Appendix, the measured inlet and outlet O_2 concentrations for A/H's A and B were used to calculate the amount of air to gas leakage. Then, the amount of leakage, the air inlet temperature, the gas inlet flow, and the corrected gas outlet temperature were used to get an uncorrected gas outlet temperature and the air out flow.

D. Calculation of Air Out Flow

The air out flow for A/H C is assumed to be the same as the air in flow (which was measured), since it is assumed that A/H C did not leak. To calculate the air out flows for A/H's A and B, there are actually three methods available. The first method simply involves subtracting the calculated leakage flow from the measured air inlet flow. The second method uses a heat balance between the air and gas flows, which involves the four terminal temperatures, the gas entering flow, and an assumed specific heat ratio. The third method uses pressure drop data, fluid properties, the

4

measured gas in flow, and the assumption that the ratio of the gas flow resistance to the air flow resistance (Eu_g/Eu_a) does not change with the amount of fouling on the heat transfer surface. The first two methods both rely on the leakage calculation in order to get the uncorrected gas out temperature, while the third method depends on the assumption that the Eu_g/Eu_a ratio does not change as a result of fouling. The details of the second and third methods of calculating the air out flow are discussed in the Appendix.

When these three methods of calculating the air out flow were applied to the test data, it was found that the second and third methods agreed fairly well with each other, but that the air out flow obtained by the first method often deviated significantly from the other two. For this reason, it was decided that the air out flow to be used in the analysis should be the average of the second and third methods. In calculating the air out flow based on a heat balance, a constant value of 0.95 was used for the ratio of the air specific heat to the gas specific heat. This is a typical value for air heaters on coal fired boilers. The value of b (where b is the slope of the f vs. Re curve -- see the Appendix) that was used for A/H A was different than the value of b for A/H B, because the relative lengths of smooth and undulated heat transfer surfaces in the two A/H's differed. Both b values were between -0.3 and -0.5.

A comparison of the mass flows from all three methods is given in Figures 1 and 2 for A/H's A and B, respectively. It can be seen that there is a lot of scatter in the air flows obtained from the leakage calculation, but that both the second and third methods (heat balance and flow resistance) yield air flow values that lie within ±10% of the air flow used to calculate performance, which is just the average of the flows from those two methods.

E. Calculation of ε, C*, and Ntu

The Appendix gives equations for finding the effectiveness (ε) and the capacity ratio (C*) from just the four terminal temperatures. From the effectiveness and the capacity ratio, the value of Ntu was found. Figures 3-5 show the trend of effectiveness over time for all three A/H's, and Figures 6-8 show the trend of Ntu with time.

F. Calculation of Flow Resistances (see Appendix)

Values of Eu_g for all three A/H's and Eu_a for A/H's A and B were calculated by simply dividing the measured pressure drops by the calculated velocity heads (velocity head $= \rho V^2 = G^2/\rho$, where $G = \rho V$). For the air sides of A/H's A and B, the calculated air out flows were used. Figures 3-5 show the measured ΔP's for all three A/H's over time, and Figures 9-13 show the trend of Eu with time.

IV. Discussion of Results

A. A/H A (2 layer regenerator) - See Figures 3, 6, 9, & 12

☐ The effectiveness and gas side pressure drop as a function of time are shown in Figure 3. The effectiveness varied from a low of around 85-86% in Jun-Aug. of 1994 (months 14-16) up to a maximum of 91% in late 1993 (months 5-8). There is only a slight degradation in effectiveness from the beginning of the test to the end, from about 88% down to about 86%. There is

5

a noticeable dip in effectiveness that coincides with the period of high pressure drop in months 14-16.

☐ The gas side pressure drop (ΔP_g) for A/H A shows definite effects of fouling and probably some plugging. The first six months of operation show a relatively constant value of ΔP_g, but then a steady increase from about 4 in. WG to over 9 in. WG during the next 6 months. Water washing A/H A in May 1994 brought ΔP_g down again to the 5.5-7.5 range during month 14, but an apparent sootblower failure allowed ΔP_g to quickly jump up to 13 in. WG in month 15. Rewashing of A/H A cut ΔP_g down to about 6 in. WG, and then another sporadic increase up to 9 in. WG occurred during months 16-20. The final 3-4 months of operation saw another increase in ΔP_g from about 5.5 up to 7 in. WG. A plot of the gas side Eu numbers as a function of time is shown in Fig. 9. They follow the same trend as the gas side ΔP.

☐ The air side ΔP (Fig. 3) for A/H A follows roughly the same trends as the gas side ΔP, except that the ΔP_a values are lower due to the lower velocity head on the air side. A plot of the air side Eu numbers as a function of time is given in Fig. 12. They follow the same trend as the air side pressure drops, and they correlate well with the gas side Eu numbers, as expected.

☐ A plot of Ntu vs. time for A/H A is given in Figure 6. The value of Ntu is a somewhat better indicator of the heat transfer capability of the A/H than ε is, because ε is a stronger function of the air flow rate. The value of Ntu is somewhat dependent on the air and gas flow rates through the A/H, but it should reflect to a large degree the amount of fouling on the heat transfer surfaces of the A/H. The Ntu values for A/H A show a greater loss of performance than the ε values. Fig. 6 shows that there is a general decline in Ntu from an initial value of about 3.75 down to a final value of about 3.3. Also, the period of very high pressure drops in month 15 coincides with a dip in Ntu down to about 3.1.

B. A/H B (3 layer rotary regenerator) - See Figures 4, 7, 10, & 13

☐ The effectiveness and gas side pressure drop as a function of time are shown in Figure 4. The effectiveness reached a low of around 81-82% in Jun-Aug. of 1994 (months 14-16) and a high of 91-92% in Dec. 1993 (month 8). As with A/H A, A/H B did not show any significant decrease in effectiveness over time, with the possible exception of the ε values from months 25-26. The data for these last two months seems to be questionable, however, because there is no known reason for the low ε values. The low ε values were the result of air out temperature readings that were about 40 °F lower than the earlier data, possibly as a result of an instrumentation problem with the thermocouples in the air out duct. It is suspicious that these ε readings are so low immediately after the A/H had been washed, and at a time when the pressure drops are relatively low. Therefore, these data points are labeled as "questionable" on Figure 7, and are omitted from Table 2.

☐ The gas side pressure drop for A/H B shows some signs of fouling and possibly some plugging. The first six months of operation show a relatively constant value of ΔP_g followed by a spike up to 10.5 in. WG in month 7. This is followed by a sporadic drop down to almost 5 in. WG by the end of month

6

12. Water washing of the A/H in month 13 lowered ΔP_g further, down to about 3.5 in. WG. A general upward trend in ΔP_g during months 14-19 is culminated by severe spikes in months 20 and 21. A water washing in early 1995 brought the ΔP_g back down to 4-5 in. WG in months 25-26. A plot of the gas side Eu numbers is given in Fig. 10. They follow the same general trend as the gas side ΔP, but with somewhat less scatter in the data.

☐ As with A/H A, the air side ΔP for A/H B also follows the same trends as the gas side ΔP. The air side Eu numbers vs. time for A/H B are presented in Fig. 13. They follow the same general pattern as the air side pressure drops shown in Fig. 4, and they agree quite closely with the gas side Eu numbers shown in Fig. 10.

☐ A plot of Ntu vs. time for A/H B is given in Figure 7. Although the Ntu values for A/H B show the same general trend as the Ntu values for A/H A, the drop in Ntu over the period of the test is smaller than it is for A/H A, dropping from about 3.5 to about 3.25 during the first 21 months of the test period (ignoring the questionable data in months 25-26). This is attributable to the fact that A/H B received much less NH_3 than did A/H A.

C. A/H C (Heat pipe recuperator) - See Figures 5, 8, & 11

☐ The effectiveness and ΔP_g of A/H C are shown in Figure 5. The effectiveness trend for A/H C is different than it is for A and B, with a general decrease in ε over the first 19 months of testing from about 74-75% down to about 63%. However, during months 20 and 21, there was a partial recovery in ε back up to around 68-73%. Final ε values for A/H C were around 64% in months 25-26.

☐ The ΔP_g of A/H C is fairly constant during months 1-5, and then it gradually increases in months 6-8. An apparent cleaning reduced ΔP_g back down to less than 2 in. WG, but then a gradual increase over the next 9 months raised it to around 3 in. WG. A sharp spike up to 14 in. WG, a ΔP_g reduction, and then another increase to about 12 in. WG occurred in months 19-21. The final ΔP_g values for A/H C ended up at 6-7 in. WG in months 25-26. The gas side Eu numbers vs. time for A/H C are shown in Fig. 11. They follow the same trends as the gas side ΔP's.

☐ No measurements of ΔP_a were made for A/H C.

☐ A plot of Ntu vs. time for A/H C is given in Figure 8. Although the Ntu values for A/H C show the same general trend as the ε values, the value of Ntu is seen to drop from an initial value of about 2.40 down to a value of about 1.40 at the end of month 19. A recovery of Ntu back up to about 1.8 occurred in month 20. Values of Ntu for A/H C during months 25-26 were not available due to insufficient flow data. There seemed to be little correlation between the Ntu decreases and the pressure drop increases for A/H C, except that during months 11-18 the steady drop in Ntu coincided with a steady rise in ΔP_a.

D. Measured NH_3 slips into the A/H's

Each of the three A/H's was located downstream of its own SCR reactor, and each reactor contained its own particular type of catalyst. The operation and performance of the different catalysts determined how much NH_3 slip entered each of the three

7

A/H's. Periodic measurements of the NH_3 slip indicated that A/H's A and C consistently received higher concentrations of NH_3 than did A/H B by a factor of about 4.5. A tabulation of estimated NH_3 slips into each A/H is given in Table 1.

V. Conclusions

As measured by the drop in Ntu from the initial values to the final values, the thermal performance of A/H's A, B, and C dropped by about 11%, 7%, and 25%, respectively. The fact that A/H B seemed to deteriorate less than the others is attributed to the fact that A/H B received significantly less NH_3 slip than the other two A/H's (see Table 1). It is believed that A/H C deteriorated the most because it is a recuperator, and is therefore more sensitive to a given amount of fouling than a regenerator.

As one would expect, the gas side pressure drops were more sensitive to the degree of fouling and plugging than the Ntu values were. In general, all three A/H's showed generally steady increases in ΔP_g during the test period, punctuated by occasional spikes which may have been caused by a system upsets such as a sootblower failure.

In general, the high ΔP's could be reduced by aggressive cleaning methods, including sootblowing at 4 hour intervals, thorough water washing, and occasional increases in the gas outlet temperature. It was not possible, however, to maintain the original, clean ΔP of any of the A/H's. The air and gas Eu numbers for A/H A increased by 145% and 115%, respectively, from the beginning of the test to the end. For A/H B, the increases in both Eu numbers were in the 50-55% range.

Although the 3 layer A/H appeared to perform better than the 2 layer A/H, it cannot be concluded that the 3 layer design is superior to the 2 layer design. This is because the 2 layer A/H received much more NH_3 than the 3 layer A/H -- possibly as much as four to five times more. Given this significant difference in operating conditions, the 2 layer A/H performed remarkably well, and might very well have done better than the 3 layer design if the concentration of NH_3 into the two regenerators had been equal.

The Q-Pipe® A/H seemed to steadily lose thermal performance with time during most of the test, although a partial recovery was achieved during months 19-20. It is possible that some part of this performance loss may have been due to the loss of some of the heat pipes in the unit as a result of the sootblower jet penetrating the tube wall and the consequent loss of heat transfer fluid from those pipes. However, the majority of the drop in Ntu is believed to be attributable to fouling of the finned tubes on the gas side.

A separate report (see Reference 1) dealing with the corrosion tests on various heat transfer surface materials used in A/H's A and B recommends that enameled heat transfer surface should be used for Ljungstrom® air heaters when ammonia and sulfur compounds are both present in the gas stream.

Table 2 summarizes the changes in Ntu, Eu_g, and Eu_a for the three A/H's. It compares worst case values and final values to initial values of all three parameters based on Figures 6-13. Worst case increases in Eu_g varied from 185% up to 320%. Worst case increases in ΔP_a varied from 245% up to 345%. Final air and gas Eu numbers were about 115-145% higher than initial values on A/H A, and about

50-55% higher on A/H B. This is roughly consistent with the declines in Ntu (final vs. initial) of 11% for A/H A and only 7% for A/H B, since the A/H with the higher increase in Eu (A/H A) also had the greater loss of Ntu..

References

1. Harting, S. F. "Corrosion Testing Results Ljungstrom® & Q-Pipe® Air Preheaters Conducted at Gulf Power Plant Crist", ABB Air Preheater, Inc. Engineering Report 96-1, Feb. 19, 1996

2. Hinton, W. S., Personal Communication, Jan. 16, 1995

Table 1: Estimated NH$_3$ Slips into Each A/H

a. A/H A

Dates	No. of Days of Operation	NH$_3$ Conc. (ppm)	Days * Conc. (ppm-d)
9/23/93-1/6/94	129	5.7	735.3
1/7/94-6/24/94	74	2	148
6/27/94-10/24/94	99	2.2	217.8
10/25/94-6/3/95	88	3	264
6/4/95-7/10/95	38	3.3	125.4
Total	428	——	1490.5
Average	——	——	3.5

b. A/H B

Dates	No. of Days of Operation	NH$_3$ Conc. (ppm)	Days * Conc. (ppm-d)
9/23/93-12/15/93	63	0.9	56.7
12/15/93-1/7/94	12	1.2	14.4
1/7/94-7/12/94	135	0.4	54
7/13/94-10/25/94	92	0.8	73.6
10/26/94-6/3/95	82	1.1	90.2
6/4/95-7/8/95	37	0.7	25.9
Total	421	——	314.8
Average	——	——	0.75

c. A/H C

Dates	No. of Days of Operation	NH$_3$ Conc. (ppm)	Days * Conc. (ppm-d)
9/23/93-12/15/93	38	5.6	212.8
12/15/93-1/5/94	5	2	10
1/5/94-7/12/94	89	1.1	97.9
7/12/94-11/2/94	67	2.7	180.9
11/3/94-6/9/95	60	5.4	324
6/10/95-7/13/95	14	5	70
Total	273	——	895.6
Average	——	——	3.3

Table 2: Summary of Decreases in Ntu and Increases in Pressure Drops for the three A/H's
(Worst Case and Final Values Compared to Initial Values)

a. Ntu (from Figs. 6-8)

A/H	Initial Value (I)	Worst Case Value (WC)	Final Case Value (F)	% Change WC vs. I	% Change F vs. I
A	3.75	3.1	3.3	-17%	-12%
B*	3.5	2.8	3.25*	-20%	-7%
C	2.4	1.4	1.8	-42%	-25%

b. Gas Side Eu Number (from Figs. 9-11)

A/H	Initial Value (I)	Worst Case Value (WC)	Final Case Value (F)	% Change WC vs. I	% Change F vs. I
A	35	140	75	+300%	+115%
B	28	118	43	+320%	+55%
C	63	180	135	+185%	+115%

c. Air Side Eu Number (from Figs. 12-13)

A/H	Initial Value (I)	Worst Case Value (WC)	Final Case Value (F)	% Change WC vs. I	% Change F vs. I
A	40	137	98	+245%	+145%
B	35	155	53	+345%	+50%

* Excludes questionable data from months 25-26.

Fig. 1: Comparison of Air Out Flows for A/H "A"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

By Heat Bal. [m(hb)]
By Lkg. [m(lkg)]
By Flow Resis. [m(fr)]
10% Limits

Air Out Flow Calculated by Individual Means (scfm)

Air Out Flow Used to Calculate Performance = (m(hb)+m(fr))/2 (scfm)
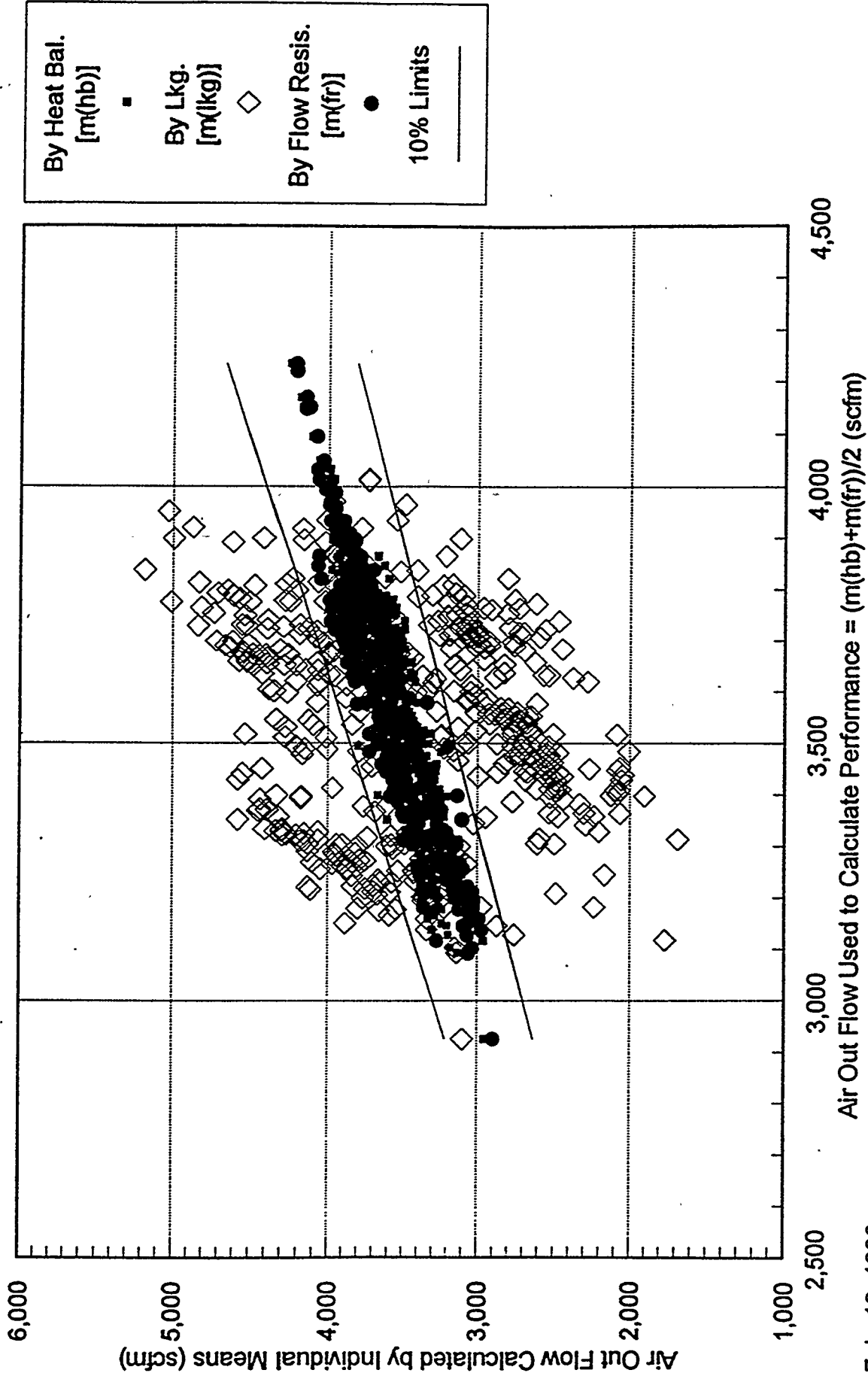
Feb. 19, 1996
C(p) Ratio = 0.95

Fig. 2: Comparison of Air Out Flows for A/H "B"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

Legend:
- By Heat Bal. [m(hb)] ■
- By Lkg. [m(lkg)] ◇
- By Flow Resis. [m(fr)] ●
- 10% Limits

Y-axis: Air Out Flow Calculated by Individual Means (scfm)

X-axis: Air Out Flow Used to Calculate Performance = (m(hb)+m(fr))/2 (scfm)
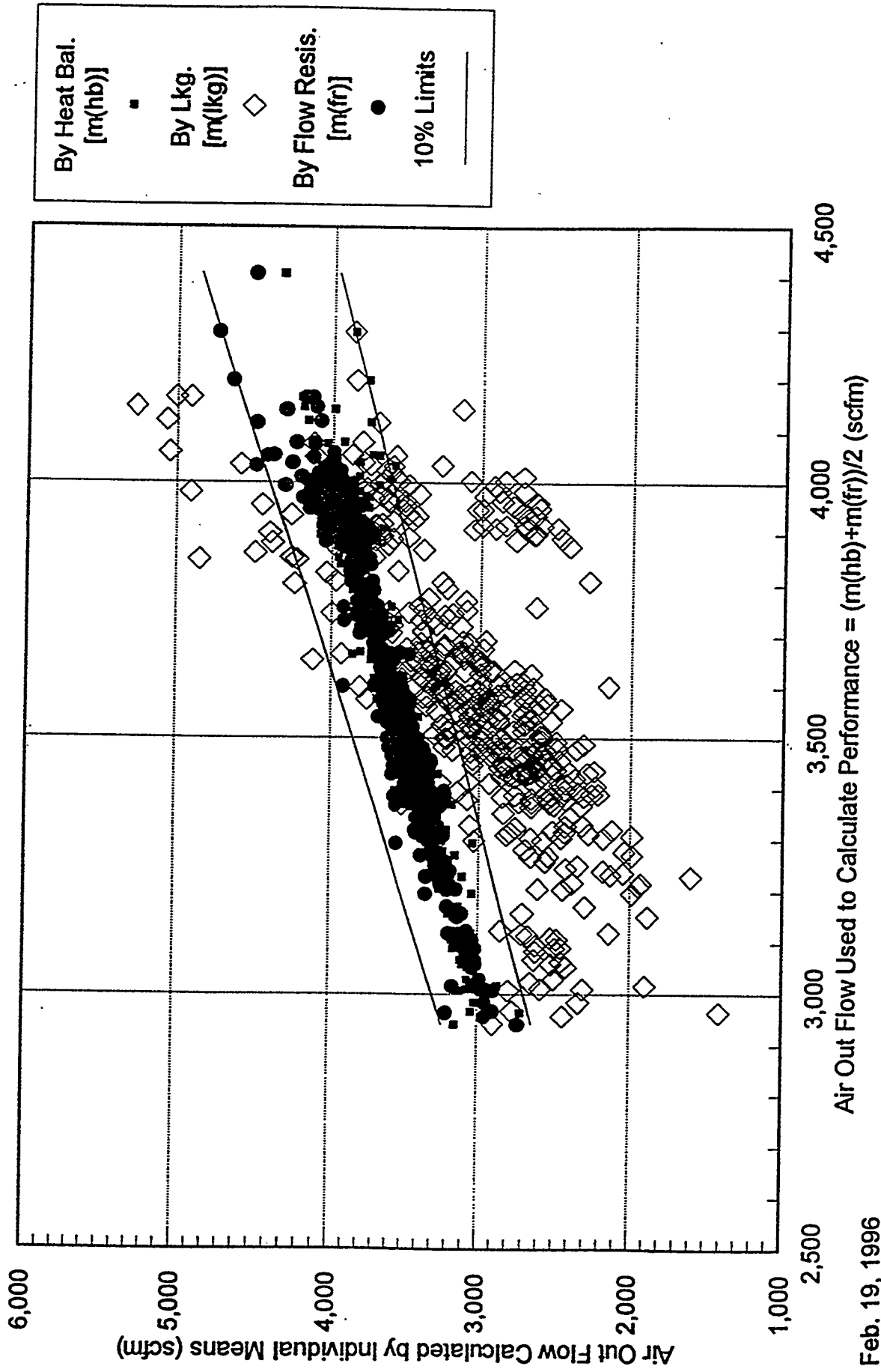
Feb. 19, 1996
C(p) Ratio = 0.95

Fig. 3: A/H "A" - Effectiveness & Pressure Drops

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
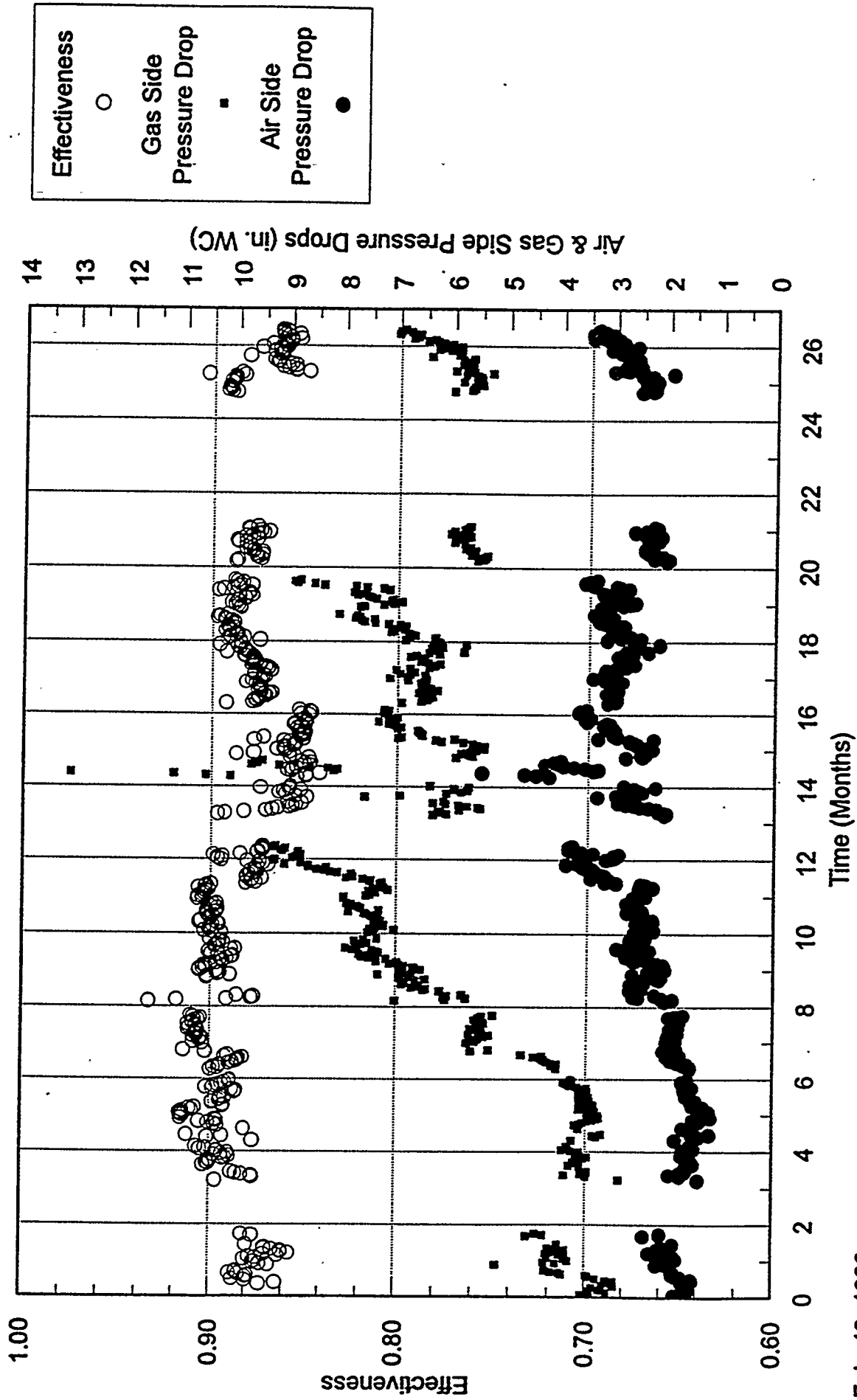
Feb. 19, 1996
A/H "A" has 2 layers of element

Fig. 4: A/H "B" - Effectiveness & Pressure Drops

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

Legend:
- ○ Effectiveness
- ▪ Gas Side Pressure Drop
- ● Air Side Pressure Drop

Air & Gas Side Pressure Drops (in. WC)

Effectiveness

Time (Months)
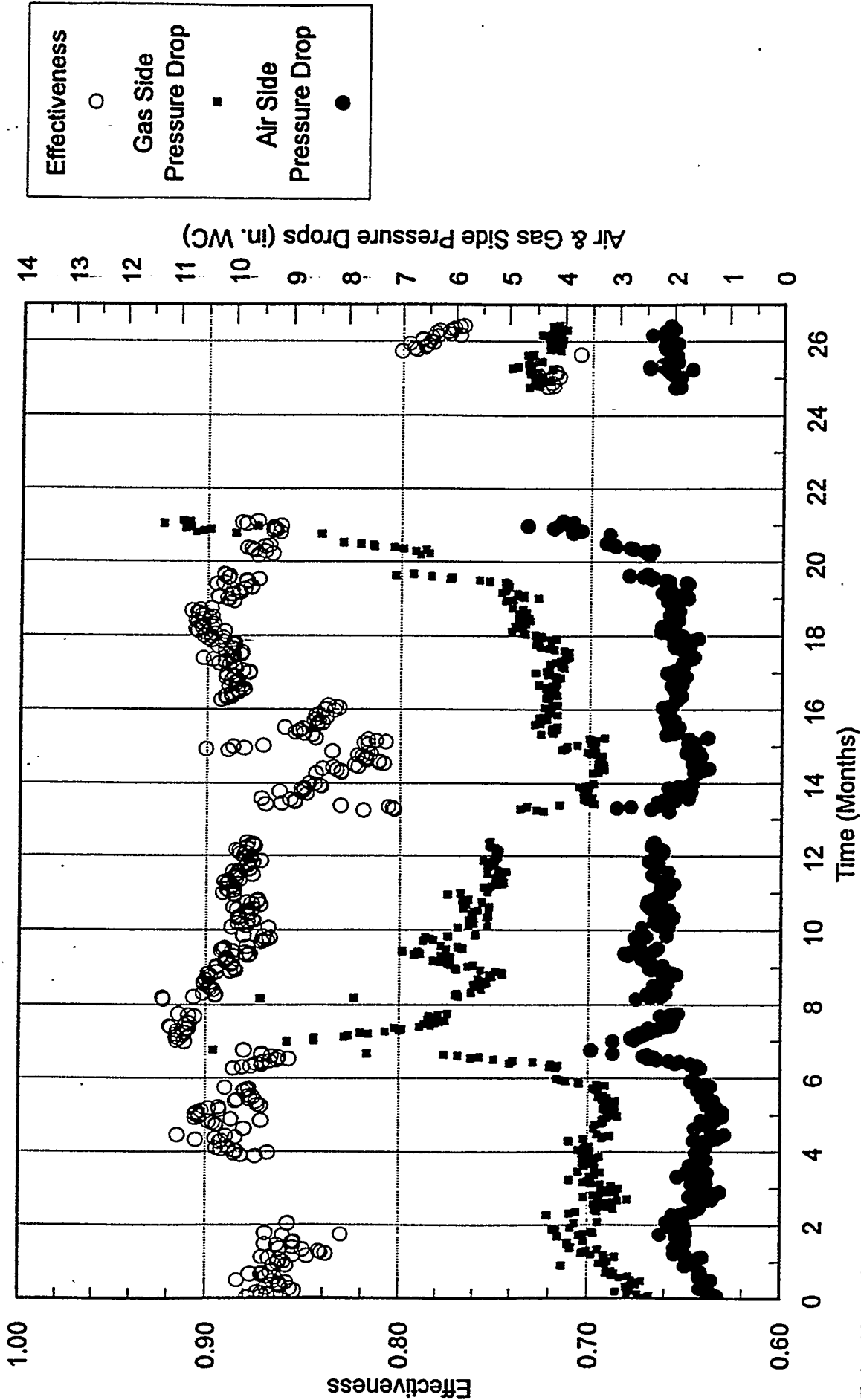
Feb. 19, 1996
A/H "B" has 3 layers of element

Fig. 5: A/H "C" - Effectiveness & Pressure Drop

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

Effectiveness	○
Gas Side Pressure Drop	■

Gas Side Pressure Drop (in. WC)

Effectiveness

Time (Months)
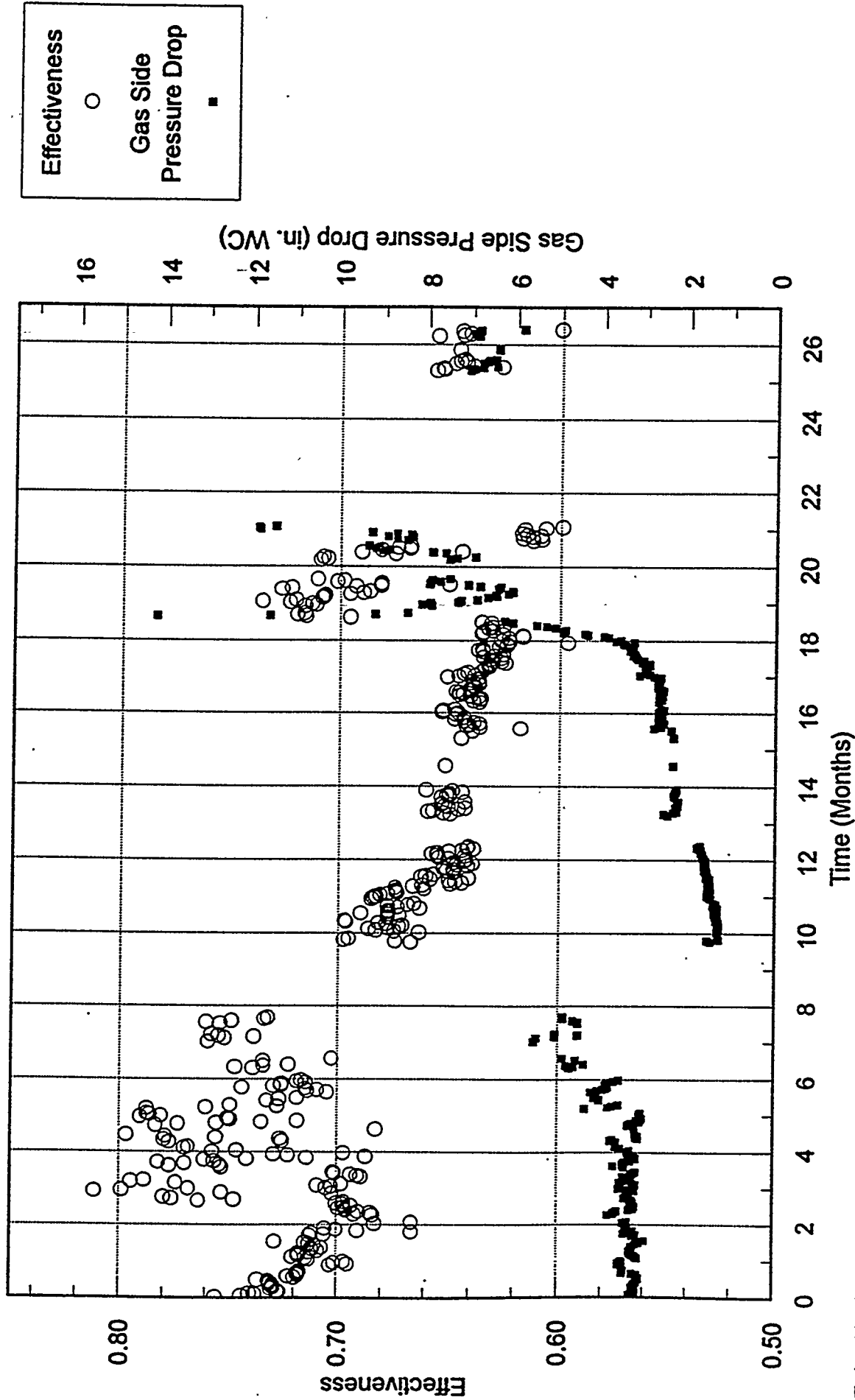
Feb. 19, 1996
A/H "C" is a heat pipe A/H

Fig. 6: Ntu for A/H "A"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
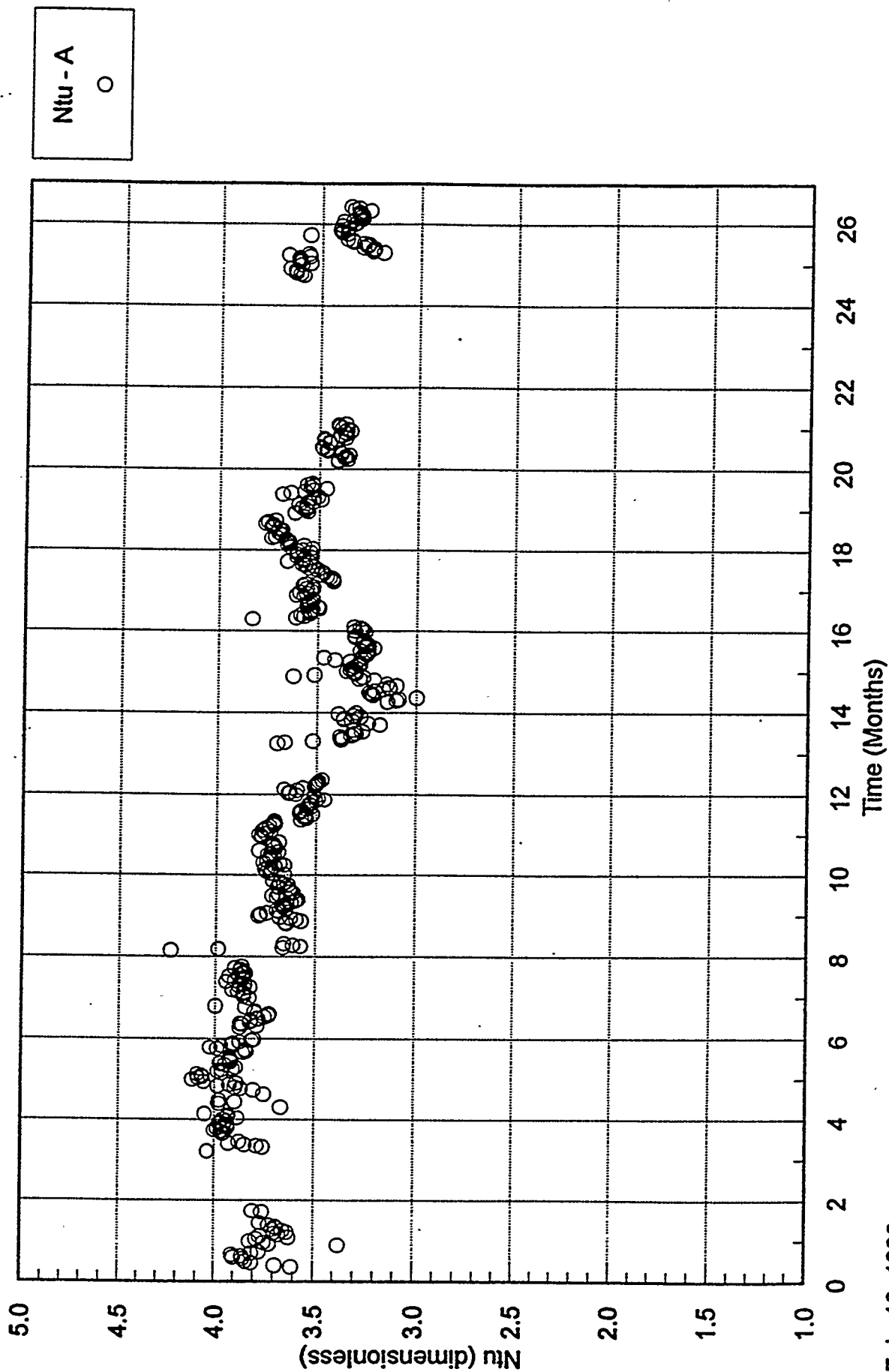
Feb. 19, 1996

Fig. 7: Ntu for A/H "B"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

Ntu - B ○   Questionable Data ◇

Ntu (dimensionless)

Time (Months)

Questionable Data
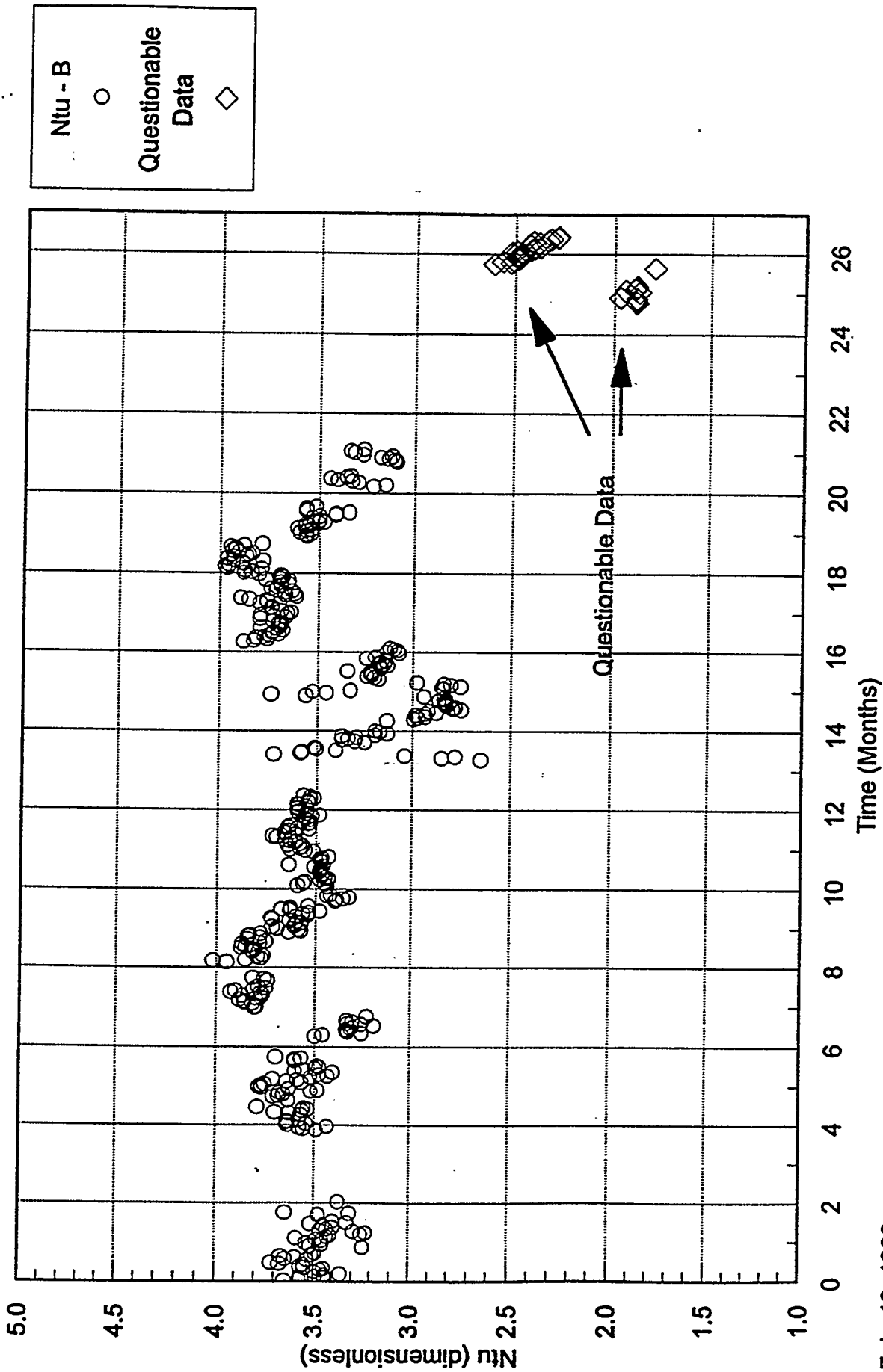
Feb. 19, 1996

Fig. 8: Ntu for A/H "C"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
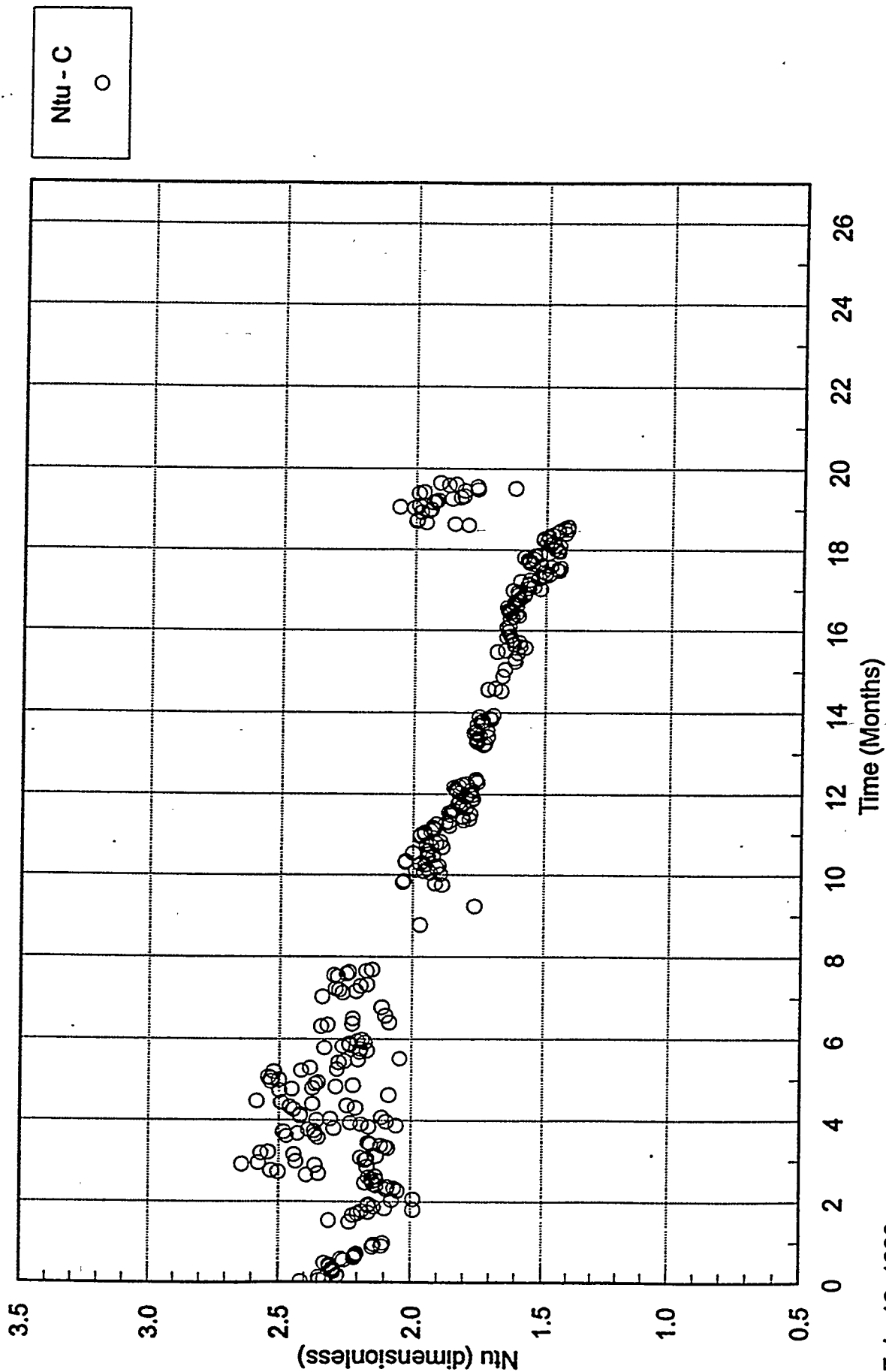
Feb. 19, 1996

Fig. 9: Gas Side Eu Numbers for A/H "A"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
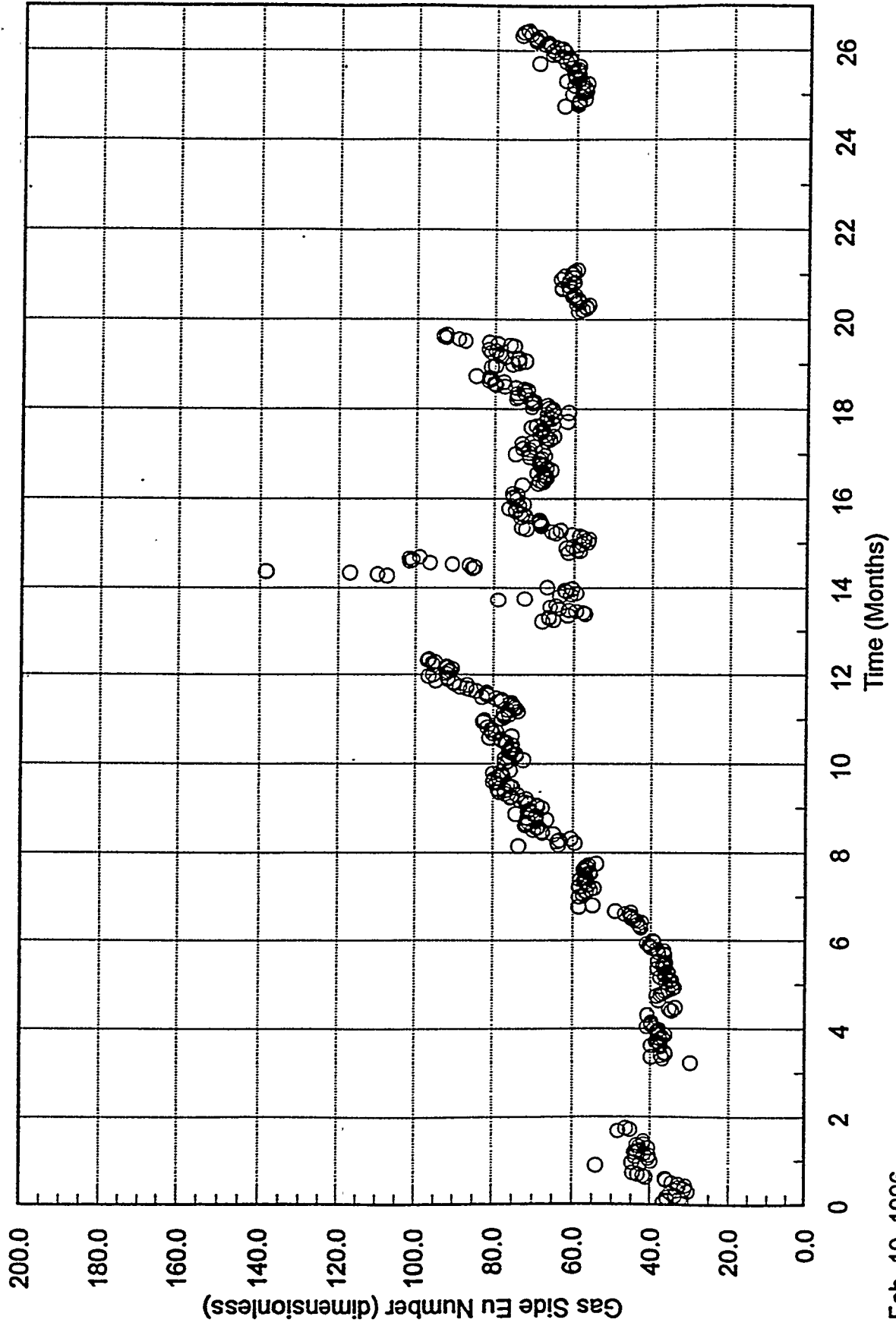
Feb. 19, 1996

Fig. 10: Gas Side Eu Numbers for A/H "B"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995

Gas Side Eu Number (dimensionless)

Time (Months)
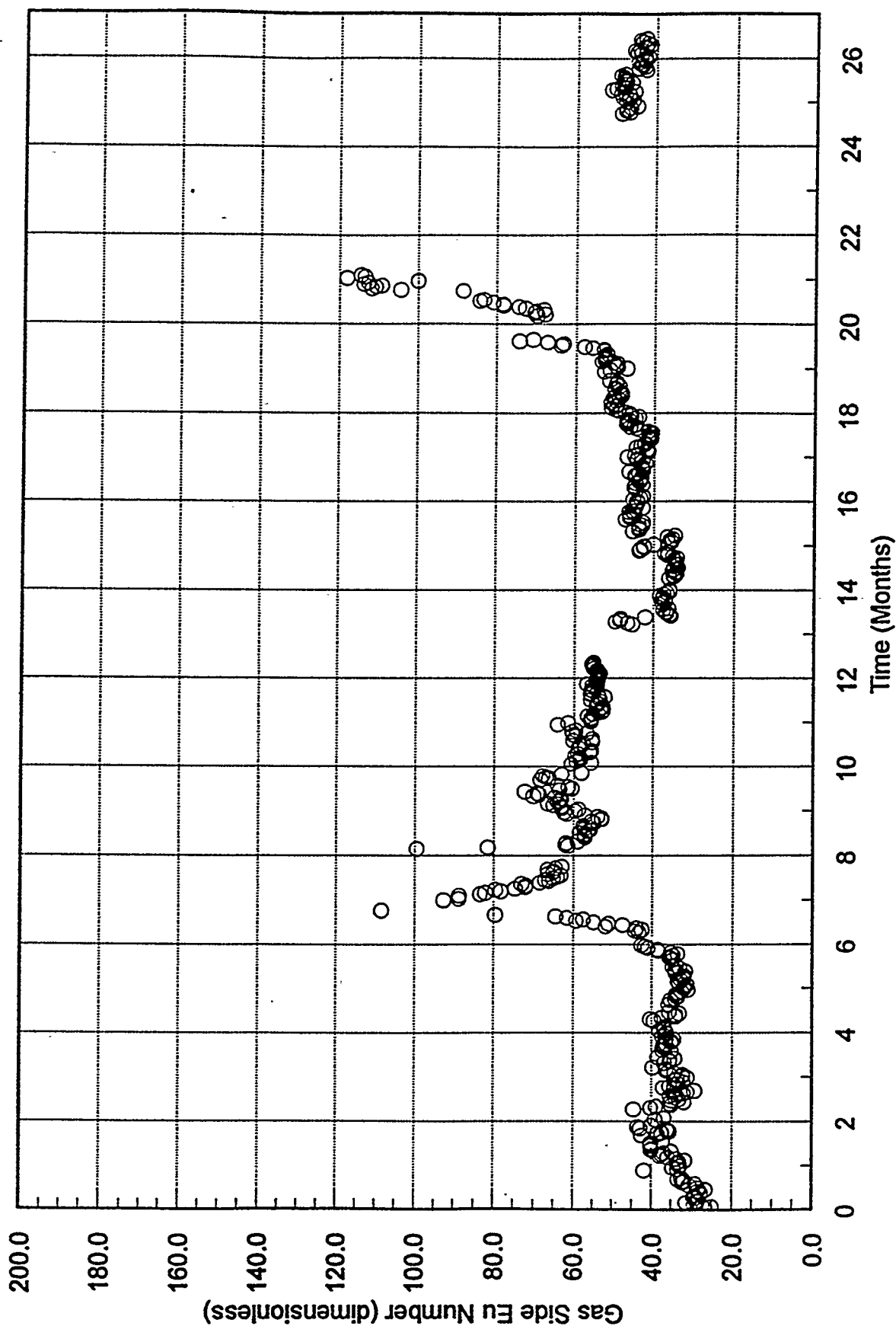
Feb. 19. 1996

Fig. 11: Gas Side Eu Numbers for A/H "C"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
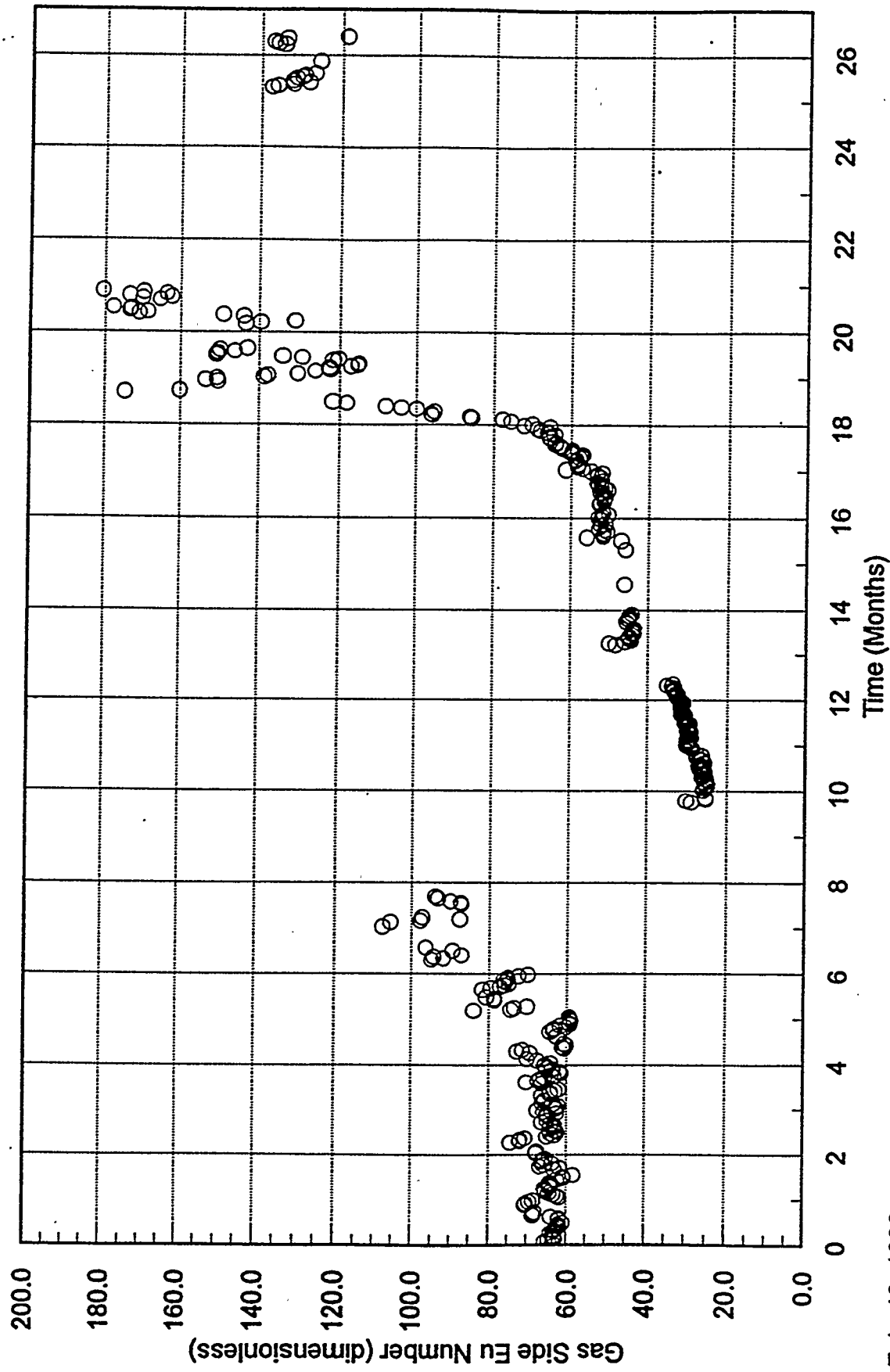
Feb. 19, 1996
Velocity based on minimum flow area

Fig. 12: Air Side Eu Numbers for A/H "A"

Southern Co. - Plant Crist #5 - May 1993 thru July 1995
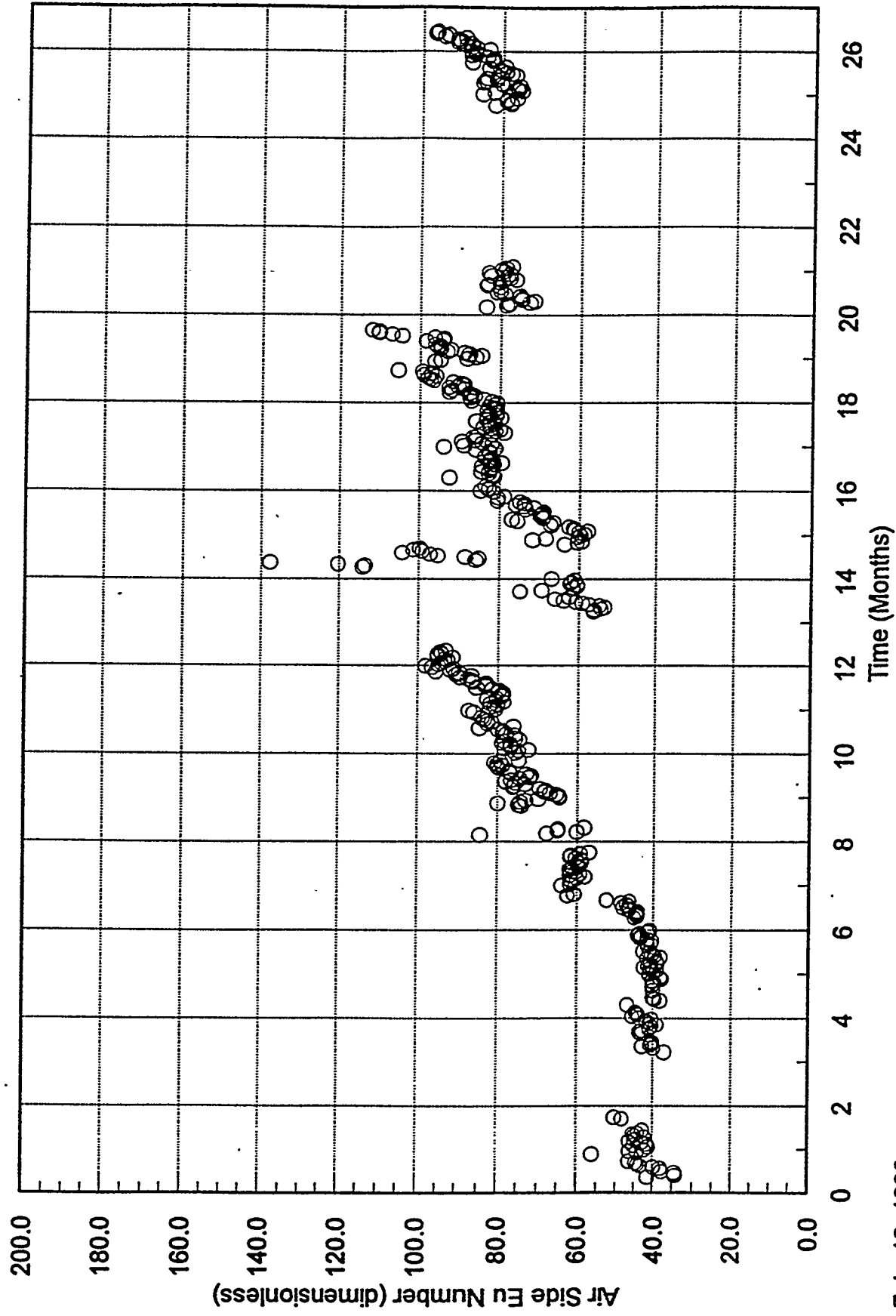
Feb. 19. 1996

Fig. 13: Air Side Eu Numbers for A/H "B"
Southern Co. - Plant Crist #5 - May 1993 thru July 1995
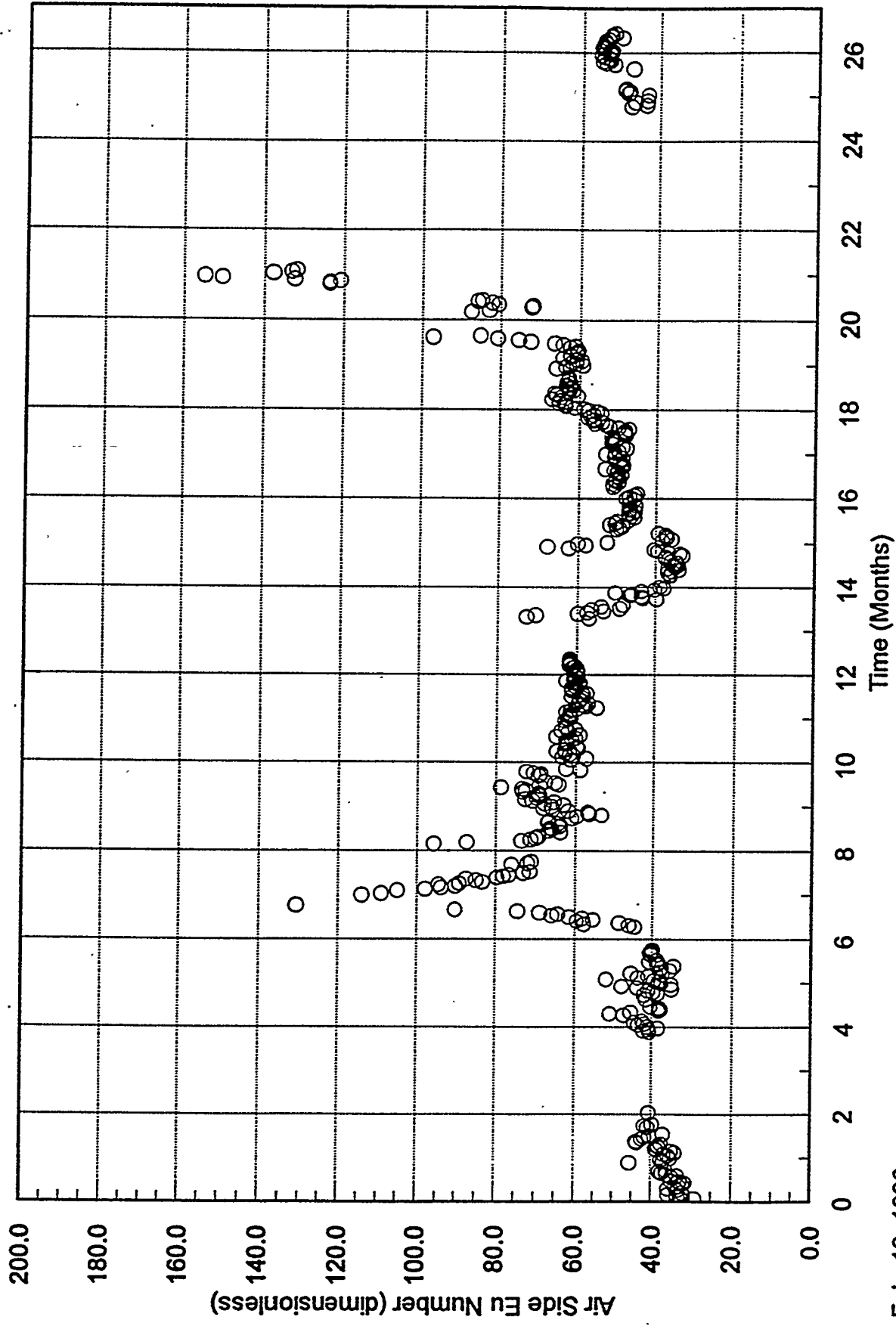
Feb. 19, 1996

Appendix

This appendix describes the nomenclature and some of the equations that were used to analyze the data from the three air heaters that were tested. Much of the material in this appendix also appears in the book <u>Compact Heat Exchangers</u> by Kays and London. The reader may wish to consult this reference for further details.

I The effectiveness-Ntu method

A. This method is normally used to calculate the two fluid outlet temperatures of a heat exchanger when the heat transfer coefficients, areas, fluid flow rates, fluid properties, and inlet temperatures are known. In this analysis, however, the outlet temperatures were measured, so values for effectiveness and Ntu were calculated and plotted as a function of time.

B. The dimensionless parameters used in this method are defined to be:

1. Effectiveness: $\varepsilon = \dfrac{q}{q_{max}} = \dfrac{C_c(T_{c,o} - T_{c,i})}{C_{min}(T_{h,i} - T_{c,i})} = \dfrac{C_h(T_{h,i} - T_{h,o})}{C_{min}(T_{h,i} - T_{c,i})}$

2. Capacity ratio: $C^{\bullet} = \dfrac{C_{min}}{C_{max}}$

3. Number of transfer units: $Ntu = \dfrac{UA}{C_{min}}$

C. In addition to the above definitions, the following nomenclature is used in this appendix:

T = Temperature (°F)
m = Mass flow (lb/hr)
c = Average fluid specific heat (Btu/lb/°F)
C = Thermal capacity (Btu/hr/°F)
C_{min} = Min(C_c,C_h) (Btu/hr/°F)
C_{max} = Max(C_c,C_h) (Btu/hr/°F)
h = Average convective heat transfer coefficient (Btu/hr/ft²/°F)
A = Surface area for heat transfer (ft²)
U = An overall convective coefficient that accounts for convection on both the hot and cold sides of the heat exchanger.
q = Heat actually transferred by the heat exchanger (Btu/hr)
q_{max} = Maximum amount of heat that could theoretically be transferred from the hot fluid to the cold fluid

Subscripts:

a - air
c - cold side fluid
c - corrected (in combination with subscripts g,o)
g - gas
h - hot side fluid
i - at inlet
o - at outlet
u - uncorrected (in combination with subscripts g,o)

D. The effectiveness of a pure counterflow heat exchanger is a function of Ntu and C^* only:

$$\varepsilon = \frac{1 - \exp(-Ntu(1 - C^*))}{1 - C^* \exp(-Ntu(1 - C^*))} \text{ for } C^* \neq 1 \quad \text{or}$$

$$\varepsilon = 1 - \exp(-Ntu) \text{ for } C^* = 1$$

Note also that if the four terminal temperatures are known, then the definition of effectiveness implies that:

$$\varepsilon = \frac{Max((T_{c,o} - T_{c,i}), (T_{h,i} - T_{h,o}))}{T_{h,i} - T_{c,i}}$$

E. If ε and C^* are known, Ntu can be calculated by solving the above equation for Ntu:

$$Ntu = \frac{1}{C^* - 1} \ln\left(\frac{1 - \varepsilon}{1 - C^* \varepsilon}\right) \text{ for } C^* \neq 1$$

$$Ntu = -\ln(1 - \varepsilon) \text{ for } C^* = 1$$

F. The value of A is usually chosen to be either $A=A_c$ or $A=A_h$. It is not important which one is selected because the value of U will change based on which A value is chosen. What is important is that the product UA be correct. For the typical Ljungstrom® rotary regenerator, $A_c=A_h=A$, and U is calculated as follows:

$$UA = \frac{1}{\dfrac{1}{h_c A_c} + \dfrac{1}{h_h A_h}}$$

Then, for the special case where $A=A_c=A_h$, this simplifies to

$$U = \left(\frac{1}{h_c} + \frac{1}{h_h}\right)^{-1}$$

II Other equations for heat transfer and pressure drop were used in the analysis

A. A heat balance on the heat exchanger requires that

$$q_c = -q_h \text{ which means that}$$

$$m_c C_c (T_{c,o} - T_{c,i}) = m_h C_h (T_{h,i} - T_{h,o})$$

This can be rewritten as:

$$\frac{m_c C_c}{m_h C_h} = \frac{(T_{h,i} - T_{h,o})}{(T_{c,o} - T_{c,i})}$$

and then we can say that

$$C^* = Min\left[\left(\frac{T_{h,i} - T_{h,o}}{T_{c,o} - T_{c,i}}\right), \left(\frac{T_{c,o} - T_{c,i}}{T_{h,i} - T_{h,o}}\right)\right]$$

Thus, simply by knowing the four terminal temperatures, both the effectiveness (ε) and the capacity ratio (C^*) can be found. If the four terminal temperatures, the specific heat ratio (c_c/c_h), and m_h are all known, the cold flow, m_c can be found as follows:

$$m_c = \left(\frac{T_{h,i} - T_{h,o}}{T_{c,o} - T_{c,i}}\right)\left(\frac{m_h}{c_c/c_h}\right)$$

B. Pressure drop (ΔP) through each side of the heat exchanger was assumed to be due to frictional losses only. Using the Fanning friction factor to relate velocity head to ΔP yields:

$$\Delta P = \frac{4fL}{D_h}\left(\frac{\rho V^2}{2g_c}\right)$$

where
  f = Fanning friction factor
  L = Flow length through heat exchanger
  D_h = Hydraulic diameter of heat transfer surface
  ρ = Fluid density
  V = Average fluid velocity
  g_c = Conversion factor = 32.2 ft-lb$_m$/(sec^2 - lb$_f$)

This equation is useful when it is applied to a Ljungstrom® rotary regenerator because the ratio L/D_h must be the same for both the cold fluid and the hot fluid, since the rotation of the heat transfer surface (from the air stream to the gas stream and back again) means that the same heat transfer surface is used on both sides, even when the heat transfer surface is fouled. Thus, the ratio of ΔP_c to ΔP_h should be:

$$\frac{\Delta P_c}{\Delta P_h} = \frac{f_c \rho_c V_c^2}{f_h \rho_h V_h^2}$$

Since ρ_c, ρ_h, V_c, and V_h can all be calculated from mass flows and temperatures, it is possible to obtain the ratio f_c/f_h:

$$\frac{f_c}{f_h} = \frac{\Delta P_c/\rho_c V_c^2}{\Delta P_h/\rho_h V_h^2}$$

C. It can furthermore be assumed that both f_c and f_h can be expressed as functions of the Reynolds number (Re) in the following forms:

$$f_c = a_c \cdot Re_c^{b_c} \qquad \text{and} \qquad f_h = a_h \cdot Re_h^{b_h}$$

and that the f vs. Re curve is the same for both fluids, so that $a_c = a_h = a$ and $b_c = b_h = b$. This assumption is known to be valid for Ljungstrom® air heater heat transfer surfaces in the clean condition, and it is expected to be true for fouled heat transfer surfaces if the fouling does not produce asymmetric deposits. An asymmetric deposit would create more drag on flow in one direction than it would on flow in the other direction (see paragraph D.3, below)
Substituting for f_c and f_h gives:

$$\left(\frac{Re_c}{Re_h}\right)^b = \frac{\Delta P_c/\rho_c V_c^2}{\Delta P_h/\rho_h V_h^2}$$

Then, since $Re = G \cdot D_h/\mu$, where G is the ratio of mass flow to flow area, and

$\rho V^2 = G^2/\rho$, we can solve for the ratio G_c/G_h as follows:

$$\frac{G_c}{G_h} = \left(\frac{\rho_c \Delta P_c \mu_c^b}{\rho_h \Delta P_h \mu_h^b}\right)^{(1/(2+b))}$$

But since the flow area on the hot side should be equal to the flow area on the cold side, it is possible to say that

$$\frac{G_c}{G_h} = \frac{m_c}{m_h},$$ so we can find m_c if everything on the right hand side

of the following equation is known:

$$m_c = m_h \cdot \left(\frac{\rho_c \Delta P_c \mu_c^b}{\rho_h \Delta P_h \mu_h^b}\right)^{(1/(2+b))}$$

Although a measured value of b for the fouled heat transfer surfaces in A/H's A and B is not known precisely, the values of b for the two heat transfer surfaces in the clean condition are -0.3 and -0.8. Thus a typical value for b might be about -0.5.

D. It is expected that fouling of the heat transfer surface could affect the ΔP of the heat exchanger in one or more of the following ways:
1. Block part of the open area that was open to flow when the surface was clean. This would increase the fluid velocity in other parts of the heat exchanger (assuming that mass flow stays constant) and therefore cause an increase in ΔP as a result of a higher average fluid velocity. Full blockage of a closed channel could have a severe effect on thermal performance as well, because it removes heat transfer area from the A/H.
2. Roughen the heat transfer surface. For pipe flow, if the ratio of the roughness height to the pipe diameter is large enough, it will raise the friction factor of the pipe. Likewise, a deposit on the A/H heat transfer surface would probably raise the friction factor simply due to its roughness. This type of fouling would probably not have a noticeable impact on thermal performance, since it does not actually remove area from the A/H.
3. Increase the form drag. Besides blocking the flow channels or roughening the surface, it is possible that the deposits could add to the ΔP by forcing the flow to move around them. This causes form drag. If the deposits are shaped asymmetrically, they could produce a different amount of form drag on the gas than they do on the air. This would mean that the ratio of flow resistances might change.

III Leakage Effects

A. All Ljungstrom® air heaters (in this case, A/H's A and B) experience leakage from the higher pressure fluid to the lower pressure fluid. In this case, some of the inlet air leaks into the gas stream, where it mixes with the gas leaving the A/H's.

B. For the purpose of analyzing the performance of a Ljungstrom® rotary regenerator, it is assumed that all of this air-to-gas leakage occurs at the cold end of the A/H. This assumption allows the performance calculation and the leakage calculation to be independent of each other, which greatly simplifies the analysis.

C. It is standard practice to calculate the performance of an A/H on the basis of zero leakage, that is, as if the leakage does not occur at all. This means that the performance of an A/H is based on the following parameters:
   1. The air <u>out</u> flow, not the air in flow
   2. The gas <u>in</u> flow, not the gas out flow
   3. The <u>uncorrected</u> gas out temperature, not the corrected gas out temperature (which is what is measured). Thus the value for $T_{h,o}$ mentioned elsewhere in this Appendix would have to be the uncorrected gas out temperature.

D. This means that we must calculate the uncorrected gas out temperature based on the corrected gas leaving temperature and the measured amount of leakage.

E. Leakage Equations and Calculations:
   1. The amount of air that leaks into the gas stream is normally expressed as a percent of the gas in flow on a weight basis. It is calculated as follows:

$$\%Lkg = 100\left(\frac{\text{Air to gas lkg. (lb/hr)}}{\text{Gas in flow (lb/hr)}}\right) = 90\frac{(O_{2,out} - O_{2,in})}{(21 - O_{2,out})}$$

where

90	= Empirical constant to convert from volumetric basis to weight basis
$O_{2,out}$	= Volumetric O_2 concentration at gas out (%)
$O_{2,in}$	= Volumetric O_2 concentration at gas inlet (%)
21	= Volumetric O_2 concentration in the leaking air (%)

   2. Assuming that the air that leaks into the gas is at its inlet temperature, and that it has a specific heat that is equal to the specific heat of the leaving gas, we can find the difference between the corrected and the uncorrected gas out temperatures from a heat balance:

$$T_{g,o,u} - T_{g,o,c} = \frac{\%Lkg}{100}(T_{g,o,c} - T_{a,i})$$

This assumes that the air is the cold fluid and the gas is the hot fluid.
This equation allows us to calculate $T_{g,o,u}$ if we know $T_{a,i}$, $T_{g,o,c}$, and the leakage.

CORROSION TESTING RESULTS
LJUNGSTROM® & Q-PIPE® AIR PREHEATERS
CONDUCTED AT GULF POWER PLANT CRIST

DOE ICCT PROJECT DE-FC22-90PC89652

LAP 4202, LAP 4203, & QAP 147

Engineering Report 96-1

February 19, 1996

By: _____

Scott F. Harting
Senior Materials Engineer

Reviewed By: _____

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INTRODUCTION

On May 18, and 19, 1994, five specially prepared test baskets were loaded into the A and B Ljungstrom® air preheaters at Plant Crist. A basket had been prepared for each layer of each air preheater. Each basket contained heat transfer surface made from carbon steel, low alloy corrosion resistant steel (LACR), and porcelain enamel coated steel. There were a total of six heat transfer surface bundles in each basket as shown in the following sketch.
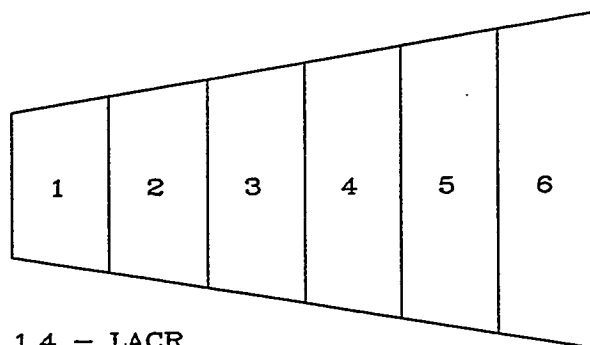

```
1,4 — LACR
2,5 — Enamel
3,6 — Carbon steel
```

Ljungstrom® air preheater A had two layers, a 30" hot layer and a 42" cold layer. Ljungstrom® air preheater B had three layers; a 18" hot layer, 42" intermediate layer, and 12" cold layer. Each of the Ljungstrom® air preheaters was on a different gas train during the testing and the ammonia slip through each gas train was not equivalent.

Although the test baskets were installed on May 18 and 19, 1994, all the baskets were subjected to an external wash on February 7, 8, and 9, 1995. Every effort was made to assure the baskets were thoroughly washed during this outage. At the end of the testing period, the baskets were removed from the Ljungstrom® air preheaters and examined. The results of this examination are reported herein.

The Q-Pipe® air preheater was examined during demolition. This unit was built as three modules so the module splits were examined. Since the heat pipes were filled with flammable liquids and solids, no plans were made to remove samples from the unit.

Conclusions

1. Ammonia bisulfate (ABS) or its corrosion products were shown to be major constituents in the Ljungstrom® air preheater deposits at temperatures less than the ABS dewpoint.

2. The magnitude of the ammonia slip had a significant effect on the corrosion losses and deposit accumulations in the units.

3. The enameled heat transfer surface was an order of magnitude superior to the LACR and carbon steel materials with regard to corrosion losses and deposit accumulations.

4. The enameled heat transfer surface exhibited superior cleanability compared to the LACR and carbon steel materials.

Recommendation

- Enameled heat transfer surface should be strongly considered for Ljungstrom® air preheaters in applications were ammonia is used for denitrification of a flue gas stream prior to the air preheater.

Results

Ljungstrom® Air Preheaters

Each pack of heat transfer surface was weighed as received from the testing then pressure washed to remove the deposits and scale. Then, each pack was weighed in the cleaned condition. Table 1 reports the original weights, the dirty weights, and the cleaned weights. Table 2 reports the percent weight lost for each pack of heat transfer surface.

Deposits were collected according to their appearance on the heat transfer surface. Differences in color, adhesion, and thickness were noted and samples were scraped from the heat transfer surface and collected. Ljungstrom® air preheater A was extensively sampled because of the ammonium slip into this unit was 3.3 ppm NH_3 and the slip into Ljungstrom® air preheater B was 0.7 ppm NH_3. This was reported in Mr. James Seebald's report, "Final Test Report on Three Air Preheaters on a SCR DeNOx Demonstration Project at Plant Crist", dated February 19, 1996. Figure 1 shows the locations of the samples.

The deposit samples from the Ljungstrom® air preheater A and one sample from the Ljungstrom® air preheater B were analyzed by combustion techniques, solution pH, x-ray fluorescence (XRF), and x-ray diffraction (XRD). These results were reported in the attached "Final Report of the Analysis of Air Heater Deposits", by Mr. Kurt Johnson, dated January 29, 1996. Table 3 summarized the x-ray diffraction results from this report.

Q-Pipe® Air Preheater

The Q-Pipe® unit was examined after removal from the test area. The overall gas inlet and outlet exhibited severe pluggage. Figures 2 and 3 show the inlet and outlet of the hot module. Figures 4 and 5 show the inlet and outlet of the intermediate module. Figures 6, 7, and 8 show the inlet and outlet of the cold module.

This unit had rotary sootblowers installed between each module (tube bundle) and water washing headers installed at outlets of the hot and intermediate layers (see figures 3 and 5). There was severe sootblowing damage noted on the water washing headers and the tubes at the inlet of the intermediate and cold modules (see figures 9, 10, 11, and 12). Close examination of figure 12 revealed that a heat pipe tube was perforated.

Discussion

There were four anomalies in the weight loss data presented in Table 1 and 2. Because of these anomalies, the weight loss and deposit loss data for each type of heat transfer surface was plotted. The bar charts are shown in figures 13 through 18. The data was statistically analyzed and this information is reported in Table 4. This analysis indicated that all the data is significant.

First, the deposit loss for Layers 1 and 2 of the hot layer for Ljungstrom® air preheater B is zero. No explanation was discovered, apparently there were no or minimal deposits on the material in these two layers.

Second, the weight gains for the cold layer basket, Layers 4 and 5 in Ljungstrom® air preheater B. The heat transfer surface was inspected and there was no evidence indicating a weight gain. It is puzzling that the LACR material did not exhibit a significant weight loss as this material was as rust coated as the other LACR materials.

Third is the relatively high weight loss of the cold layer, Layer 2 of Ljungstrom® air preheater B. The enameled heat transfer surface was examined and there were no

indications of the weight loss. The heat transfer surface sheets were all the same length and there was evidence of enamel on all edges of some of the sheets. In fact, the enameled surfaces in the cold layer still had a gloss after cleaning, indicating no corrosion to the enamel surface. There is a suspicion that this heat transfer surface pack was not weighed correctly when the baskets were assembled.

Fourth is the relatively high deposit loss for the cold layer, layer 2 of Ljungstrom® air preheater A. This loss can be attributed to a heat transfer surface channel that was plugged near the cold end and filled with flyash.

During the cleaning of the heat transfer surface, it was noted that the carbon steel and the LACR sheets were difficult to clean while the enamel coated sheets were much easier and less time consuming to clean. This cleanability is because the ABS does not corrode the enamel and does corrode the carbon steel and LACR. The corrosion products and the oxidation of the steel materials provide sites that anchor the scale and corrosion products to the steel's surface, making the cleaning process more difficult. Also, the enamel coated heat transfer sheets exhibited improved corrosion resistance when compared to the carbon steel and LACR.

The corrosion of the enameled heat transfer sheets occurs at the edges of the heat transfer sheets and proceeds at a much slower material loss. The steel heat transfer sheets are subject to corrosion over their entire surfaces, so that a slow corrosion rate can quickly remove the 0.025" to 0.040" of heat transfer surface material thickness. A similar amount of corrosion on just the edge of enameled heat transfer sheet is insignificant to the life of the heat transfer surface. Because the enameled surface is not attacked, the corrosion proceeds as though it were attacking an extremely "thick piece" of material. Therefore, enameled heat transfer surface can have a significantly longer life in a Ljungstrom® air preheater exposed to ammonia slip from a selective catalytic reduction (SCR) unit.
The report from Kurt Johnson indicates that the deposits are "hydrated iron and iron sulfate compounds"; one of the iron sulfate compounds is ammonio jarosite, $(NH_4)Fe_3(SO_4)_2(OH)_6$ which is a corrosion product of ABS and iron (steel). The other iron sulfur compounds (hydronium jarosite, rozenite, and bilinite) may be corrosion products of ABS or sulfur trioxide (SO_3). The iron oxides (hematite, maghemite, and geothite) are corrosion products of iron, oxygen, and water.

The quantity of the ammonio jarosite may be calculated from the nitrogen content reported in the combustion results.

	Percent
Sample #	Ammonio jarosite
4	10.3 %
5	27.4 %
6	13.7 %
10	44.5 %

Since ammonio jarosite is a corrosion product of ABS and steel, the ammonio jarosite in the deposits indicate that ABS has been corroding the heat transfer surface. The pressure drop fluctuations experienced by the units were attributed to ABS deposition and this deposit confirms the source of the problems.

The dewpoint of ABS is a function of the NH_3 content and the SO_3 content of the flue gas. For Ljungstrom® air preheater A the ABS dewpoint is in the range of 417°F to 426°F (213.9°C to 218.9°C) and for Ljungstrom® air preheater B the ABS dewpoint is in the range of 408°F to 416°F (208.9° to 213.3°C). The dewpoints are based on an SO_2 content that varied between 700 ppm to 1600 ppm and a 2% total SO_2 to SO_3 conversion factor. It is well known that when a material condenses from a gas, that the bulk of the condensation occurs at temperatures lower than the dewpoint. Therefore, the ABS will condense over a range of temperatures and a length of heat transfer surface. Also, the flow in the Ljungstrom® air preheater determines metal temperatures and changes the temperature profile through the unit (see Figures 19 and 20).

The percentages of ammonio jarosite present in the deposits correspond to the temperature profiles through the Ljungstrom® air preheaters. Compare Figures 1, 19, and 20 with the percent ammonio jarosite table. The greatest quantities of ammonio jarosite are found below the ABS dewpoint ranges, consistent with ABS condensation theory. Remember that the liquid ABS will attract flyash and will become a solid although the metal temperature is above the solidification temperature of the ABS. This phenomenon slows the corrosion of the steel as the flyash neutralizes the ABS and confines it. Corrosion studies of steel samples in liquid ABS showed very high corrosion rates that have not materialized in service due to the flyash in the ABS. ABS will not condense at temperatures below its freezing point but as Figures 19 and 20 show, essentially the entire cold end heat transfer surface is above the freezing point of ABS.

The pluggage evidence at the inlet of the Q-Pipe® air preheater was intensified by the damper in the inlet piping of the unit. When the damper was closed, flyash would collect on the damper and opening the damper would dump a large quantity of flyash into the unit. The damper would have been closed during demolition and removal of the unit from the steel work. The damper was open at the lay down field so the flyash trapped by the damper probably fell into the unit adding to the pluggage. The majority of the pluggage was gray colored and appeared as flyash.

File: 93m030p

TABLE 1
WEIGHT LOSS DATA

Layer #			Pairs/Layer	Original Weight	Weight Before Washing	Weight After Washing
6	CS hole	12"	8	39 lb. 4 oz.	39 lb. 1 oz.	38 lb. 2 oz.
5	Enamel	12"	7	30 lb. 10 oz.	31 lb. 3 oz.	30 lb. 12 oz.
4	LACR	12"	7	25 lb. 15 oz.	27 lb. 8 oz.	26 lb. 5 oz.
3	CS hole	12"	8	27 lb. 9 oz.	27 lb. 14 oz.	26 lb. 15 oz.
2	Enamel	12"	6	17 lb. 10 oz.	16 lb. 13 oz.	16 lb. 10 oz.
1	LACR	12"	7.5	16 lb. 6 oz.	16 lb. 11 oz.	16 lb. 1 oz.
6	CS hole	18"	7	30 lb. 4 oz.	29 lb. 12 oz.	29 lb. 11 oz.
5	Enamel	18"	3	15 lb. 13 oz.	15 lb. 12 oz.	15 lb. 11 oz.
4	LACR	18"	5	19 lb.	18 lb. 15 oz.	18 lb. 11 oz.
3	CS hole	18"	6	20 lb.	19 lb. 15 oz.	19 lb. 13 oz.
2	Enamel	18"	6	19 lb. 10 oz.	19 lb. 9 oz.	19 lb. 9 oz.
1	LACR	18"	6	13 lb.	12 lb. 11 oz.	12 lb. 11 oz.
6	CS hole	30"	6.5	48 lb. 2 oz.	47 lb. 1 oz.	41 lb. 0 oz.
5	Enamel	30"	5	39 lb. 8 oz.	39 lb. 8 oz.	39 lb. 6 oz.
4	LACR	30"	5	31 lb. 10 oz.	31 lb. 2 oz.	28 lb. 3 oz.
3	CS hole	30"	4	22 lb. 12 oz.	21 lb. 11 oz.	20 lb. 4 oz.
2	Enamel	30"	5	24 lb. 13 oz.	24 lb. 13 oz.	24 lb. 11 oz.
1	LACR	30"	6	22 lb. 4 oz.	20 lb. 7 oz.	19 lb. 1 oz.
B or 1						
6	CS hole	42"	10	93 lb.	92 lb. 13 oz.	90 lb. 7 oz.
5	Enamel	42"	7	81 lb.	80 lb. 12 oz.	80 lb. 6 oz
4	LACR	42"	8	67 lb.	67 lb. 12 oz.	65 lb. 5 oz.
3	CS hole	42"	8	58 lb. 5 oz.	58 lb. 12 oz.	57 lb. 0 oz.
2	Enamel	42"	7	52 lb. 4 oz	52 lb.. 0 oz.	51 lb. 8 oz.
1	LACR	42"	8.5	37 lb. 11 oz.	36 lb.. 11 oz.	36 lb. 0 oz.
A or 2						
6	CS hole	42"	9	86 lb. 2 oz.	84 lb. 13 oz.	75 lb. 12 oz.
5	Enamel	42"	8	86 lb. 8 oz.	86 lb. 15 oz.	86 lb. 4 oz.
4	LACR	42"	9	76 lb.	75 lb. 14 oz.	68 lb. 10 oz.
3	CS hole	42"	8	57 lb. 15 oz.	57 lb. 10 oz.	51 lb. 12 oz.
2	Enamel	42"	8	51 lb. 13 oz.	52 lb. 14 oz.	51 lb. 9 oz.
1	LACR	42"	8	40 lb. 12 oz	41 lb. 14 oz.	36 lb. 14 oz.

TABLE 2
DEPOSIT AND CORROSION LOSS DATA

Ljungstrom® Air Preheater A

Layer #	Material	Layer Thickness	Pairs/Layer	Corrosion Loss, %	Deposit Loss, %
6	CS hole	30"	6.5	14.81	12.88
5	Enamel	30"	5	0.32	0.32
4	LACR	30"	5	10.87	9.44
3	CS hole	30"	4	10.99	6.63
2	Enamel	30"	5	0.5	0.5
1	LACR	30"	6	14.33	6.73
6	CS hole	42"	9	12.05	10.69
5	Enamel	42"	8	0.29	0.79
4	LACR	42"	9	9.7	9.56
3	CS hole	42"	8	10.68	10.2
2	Enamel	42"	8	0.48	2.48
1	LACR	42"	8	9.51	11.94

Ljungstrom® Air Preheater B

Layer #	Material	Layer Thickness	Pairs/Layer	Corrosion Loss, %	Deposit Loss, %
6	CS hole	18"	7	1.86	0.21
5	Enamel	18"	3	0.79	0.4
4	LACR	18"	5	1.64	1.32
3	CS hole	18"	6	0.94	0.63
2	Enamel	18"	6	0.32	0
1	LACR	18"	6	2.4	0
6	CS hole	42"	10	2.76	2.56
5	Enamel	42"	7	0.77	0.46
4	LACR	42"	8	2.52	3.6
3	CS hole	42"	8	2.25	2.98
2	Enamel	42"	7	1.44	0.96
1	LACR	42"	8.5	4.48	1.87
6	CS hole	12"	8	2.87	2.4
5	Enamel	12"	7	-0.41	1.4
4	LACR	12"	7	-1.45	4.32
3	CS hole	12"	8	2.27	3.36
2	Enamel	12"	6	5.67	1.12
1	LACR	12"	7.5	1.91	3.75

TABLE 3
X-RAY DIFFRACTION RESULTS COMPARED TO HEAT TRANSFER SURFACE DEPTH

Ljungstrom® Air Preheater A	Ljungstrom® Air Preheater B	Major	Minor	Trace
0" to 16"		Maghemite Hematite	Butlerite Hydronium Jarosite	Quartz Rozenite
16" to 30"		Maghemite Hematite	Hydronium Jarosite	Rozenite Butlerite
30" to 35"		Maghemite	Hematite	Hydronium Jarosite Rozenite, Butlerite
35" to 44"		Maghemite	Ammonio Jarosite Hematite	Hydronium Jarosite Rozenite, Goethite
	40" to 60"	Ammonio Jarosite	Maghemite, Goethite Hydronium Jarosite	Bilinite, Rozenite Hematite, Quartz
44" to 82"		Maghemite	Ammonio Jarosite Hydronium Jarosite	Goethite, Rozenite Bilinite Quartz, Hematite
82" to 86"		Maghemite	Rozenite	Quartz, Bilinite Hydronium Jarosite Ammonio Jarosite Goethite

TABLE 4
STATISTICAL ANALYSIS - WEIGHT LOSS & DEPOSIT LOSS DATA

	Ljungstrom® Air Preheater A Standard Deviation	Ljungstrom® Air Preheater B Standard Deviation	Both Standard Deviation
Corrosion			
Enamel	0.108	2.165	1.701
LACR	2.234	1.928	5.121
Carbon Steel	1.879	0.702	5.29
Deposit			
Enamel	0.991	0.523	0.709
LACR	2.13	1.681	3.99
Carbon Steel	2.59	1.293	4.534

Figure 1
Locations of the Deposit Samples
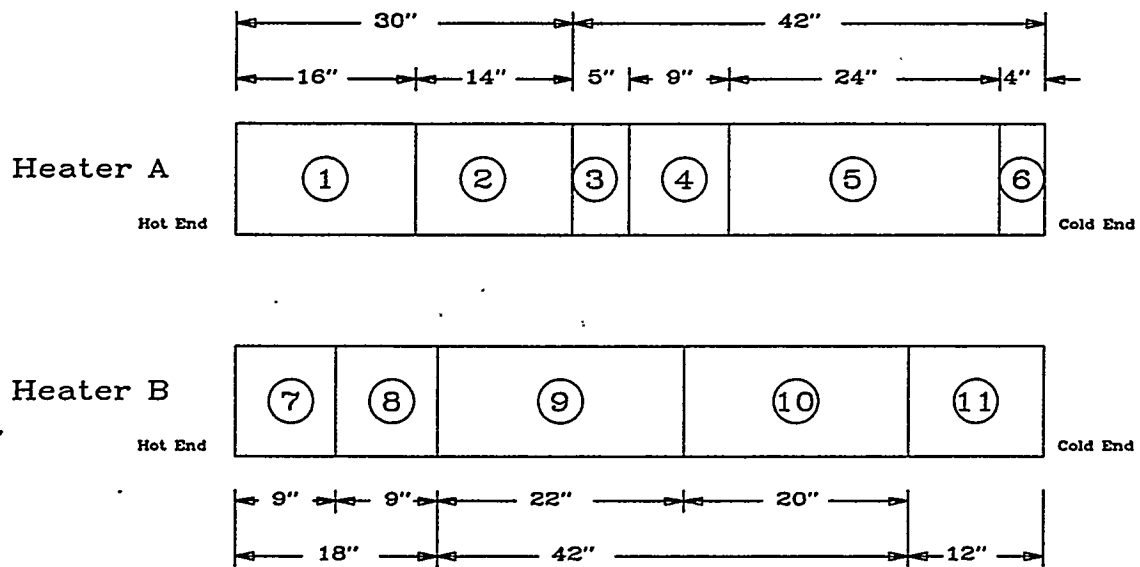
Circled numbers indicate sample numbers
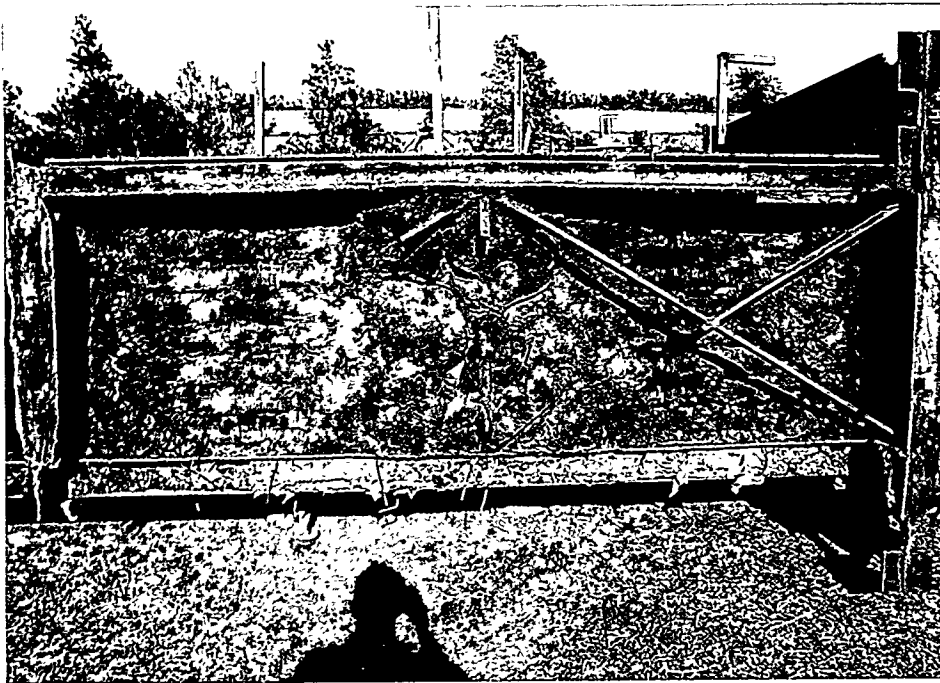
Figure 2

The gas side hot module inlet of the Q-Pipe® unit.
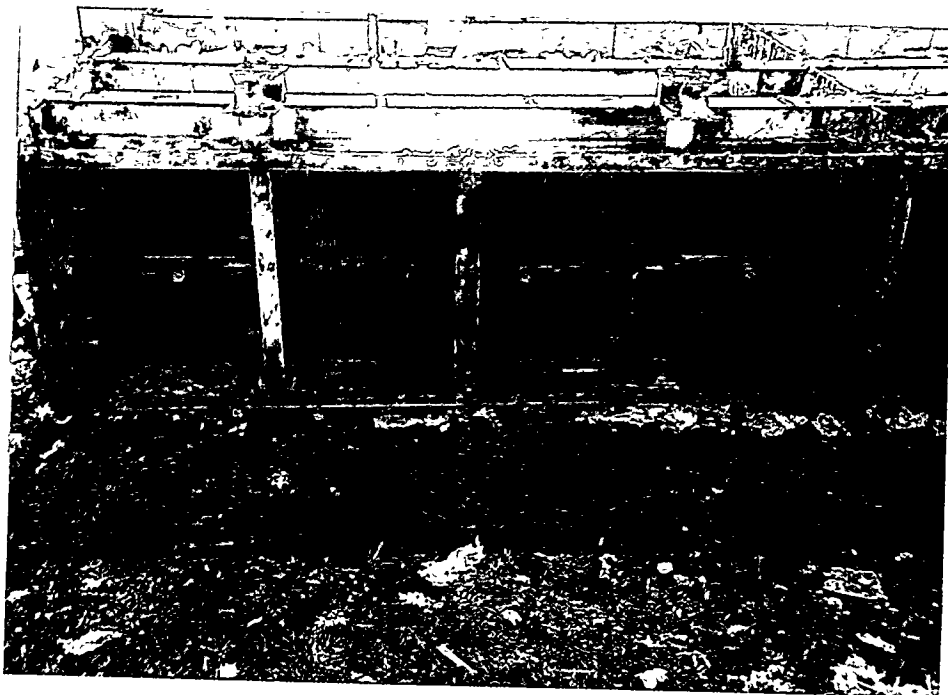
Figure 3

The gas side hot module outlet of the Q-Pipe® unit.
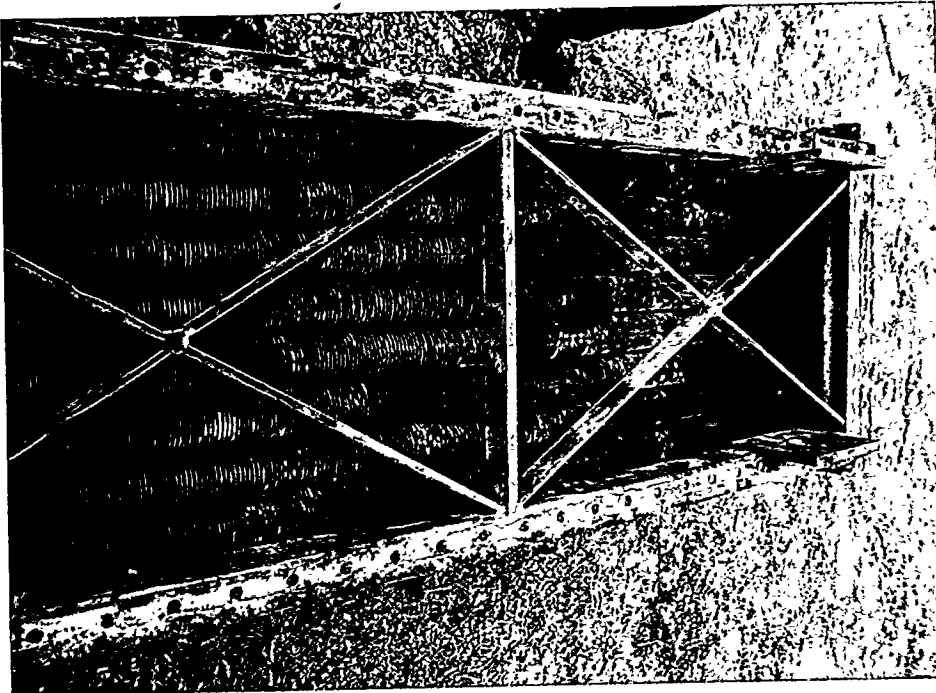
Figure 4

The gas side intermediate module inlet of the Q-Pipe® unit.
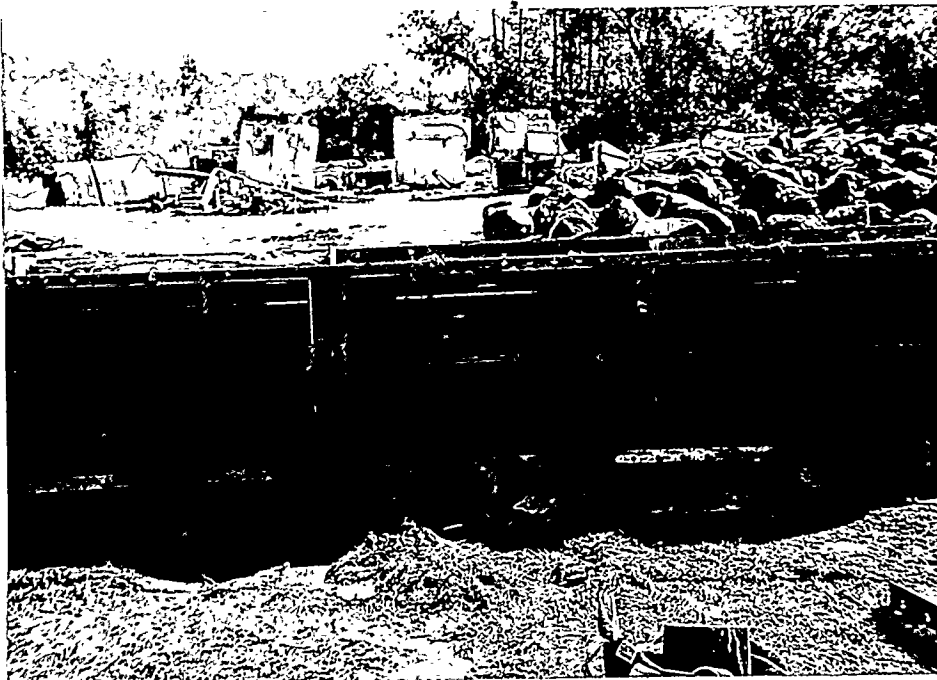
Figure 5

The gas side intermediate module outlet of the Q-Pipe® unit.
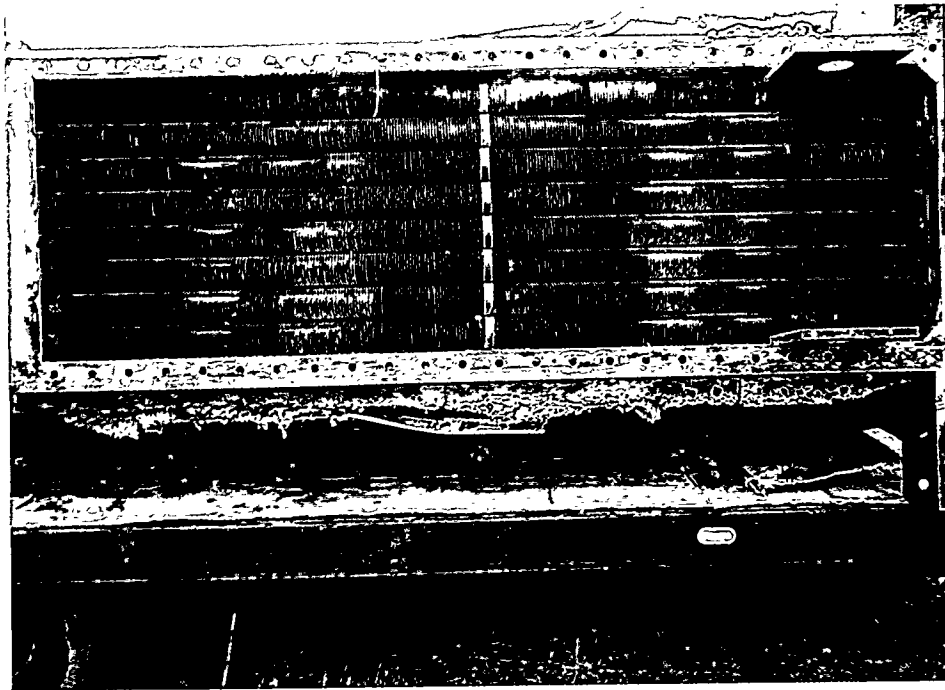
Figure 6

The gas side cold module inlet of the Q-Pipe® unit.
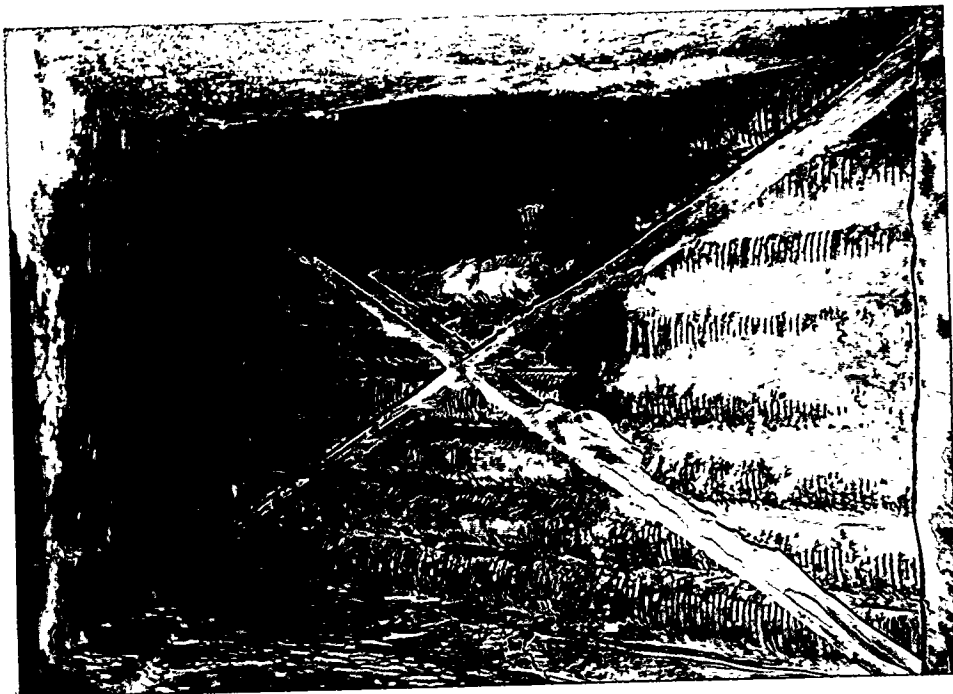
Figure 7

The gas side half of the cold module outlet of the Q-Pipe® unit.
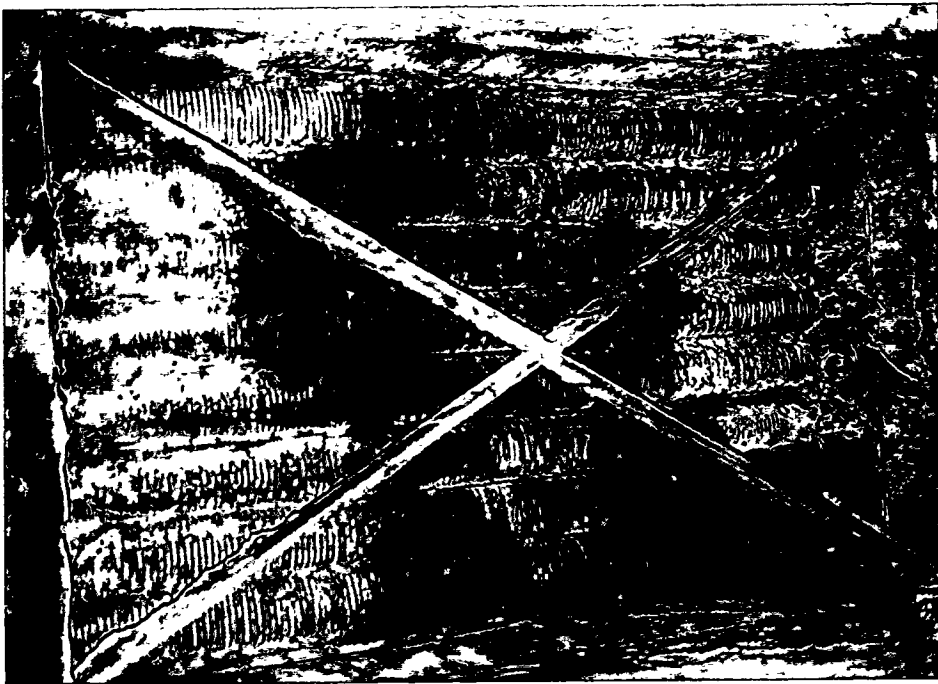
Figure 8

The gas side half of the cold module outlet of the Q-Pipe® unit.
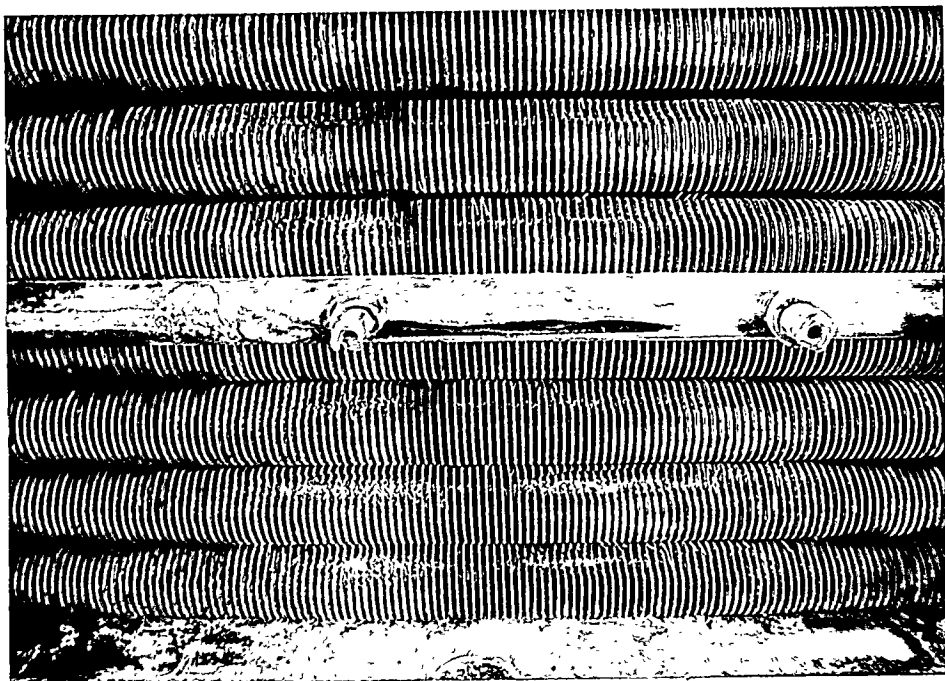
Figure 9

Sootblower damage to the water wash pipe in the outlet of the hot module.
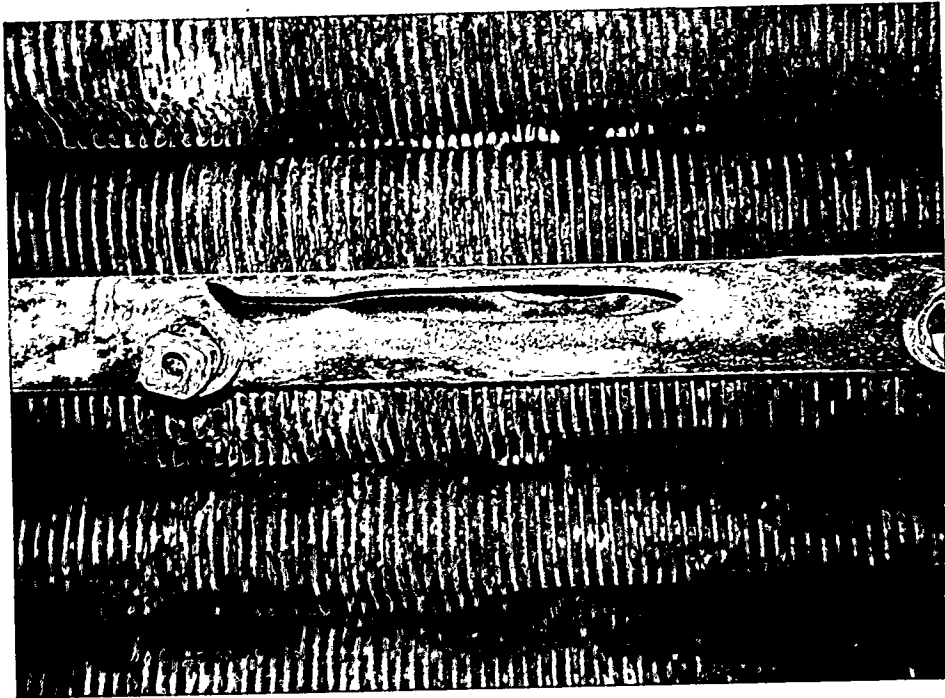
Figure 10

Sootblower damage to the water wash pipe in the outlet of the intermediate module.
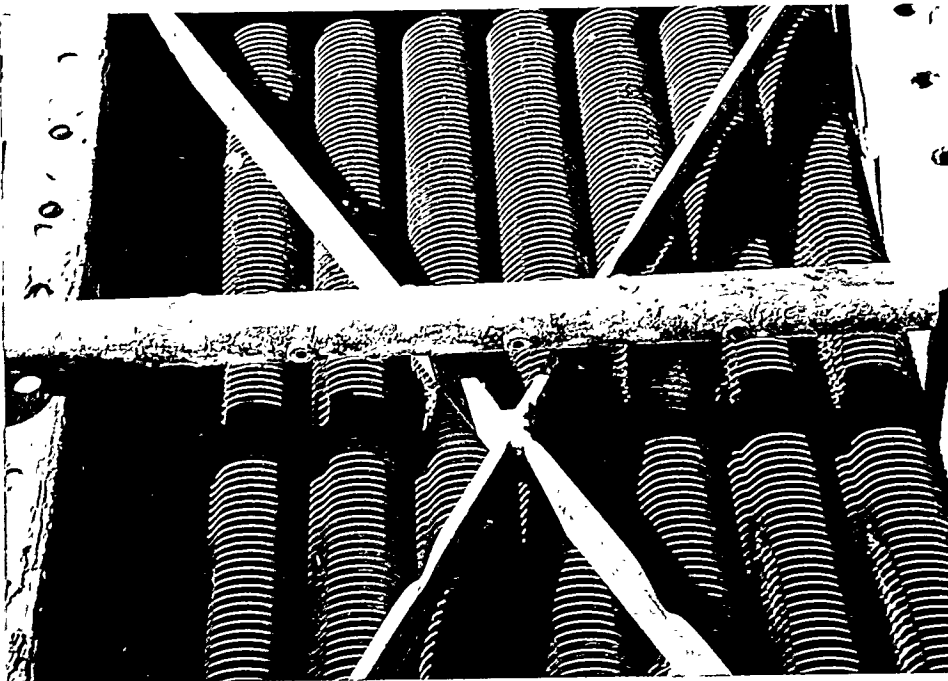
Figure 11

Sootblower damage to the heat pipes in the inlet of the intermediate module. Note that the sootblower nozzles line up with the tube and bracing bar damage.
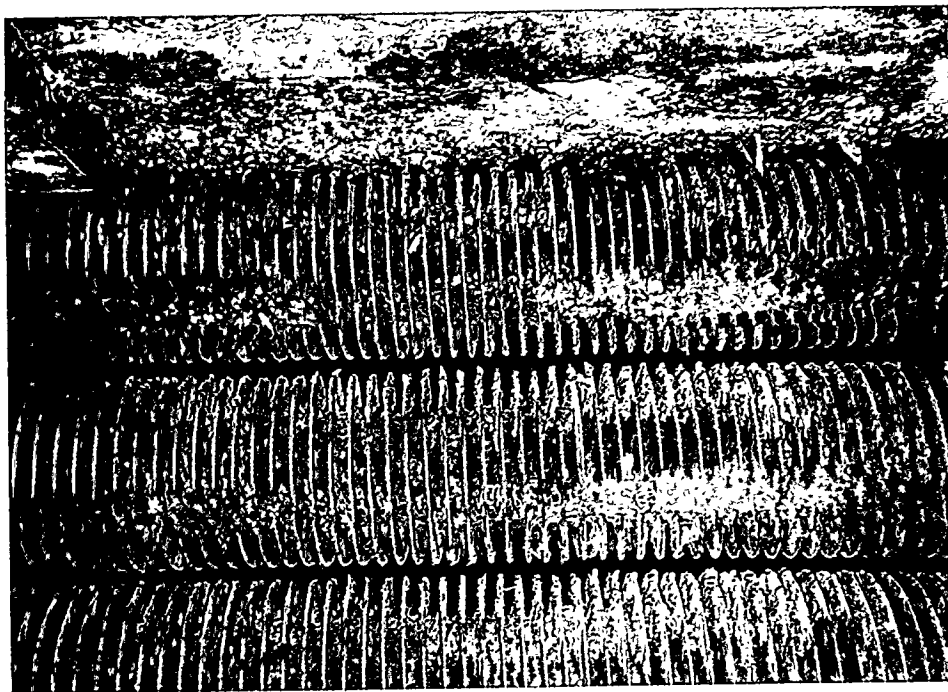
Figure 12

Sootblower damage to the heat pipes in the inlet of the cold module. Note the erosion through the tube wall in the middle tube.

Figure 13

Corrosion Loss for LACR Element

Deposit Loss for LACR Element

Figure 14

Figure 15

Corrosion Loss for Enameled Element

Figure 16

Deposit Loss for Enameled Element

Figure 17

Corrosion Loss for Carbon Steel Element

Figure 18

Deposit Loss for Carbon Steel Element
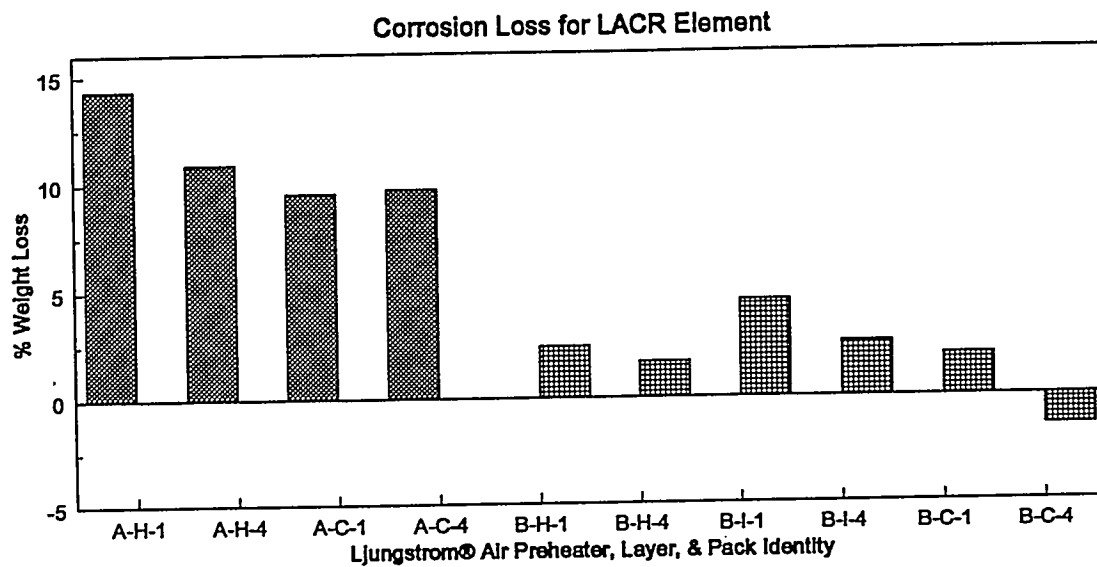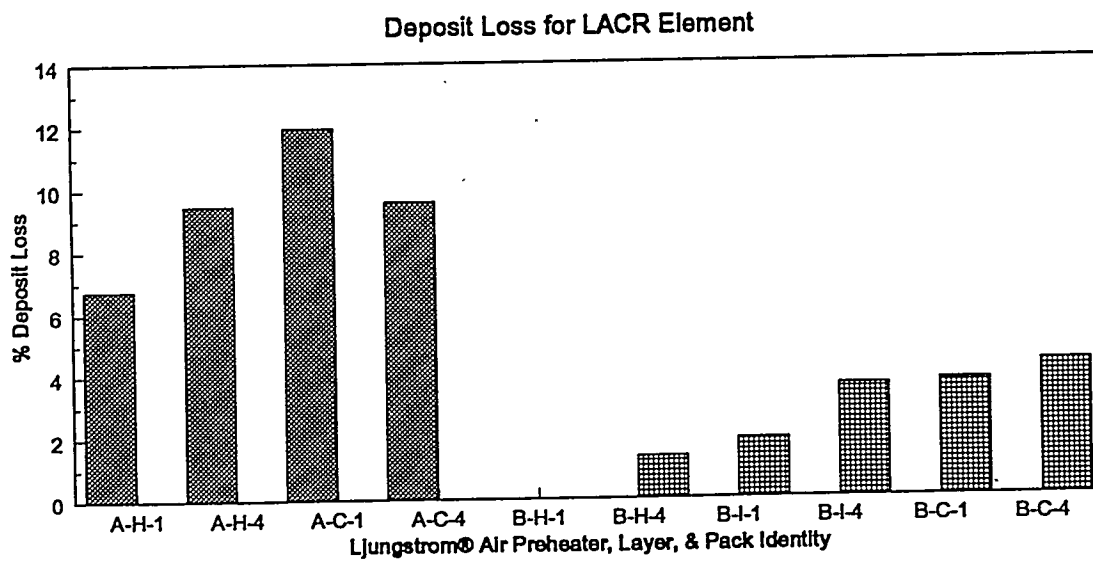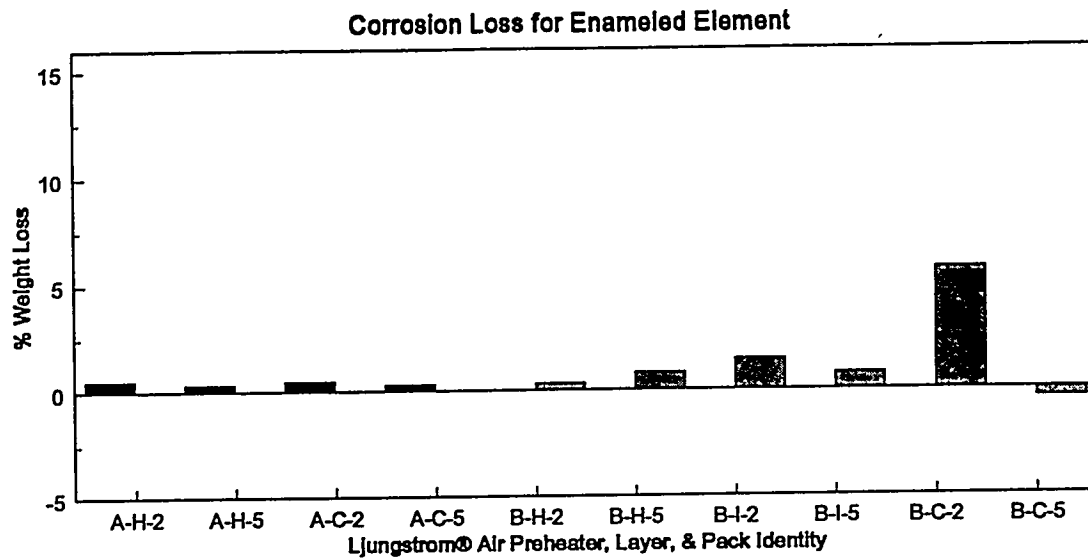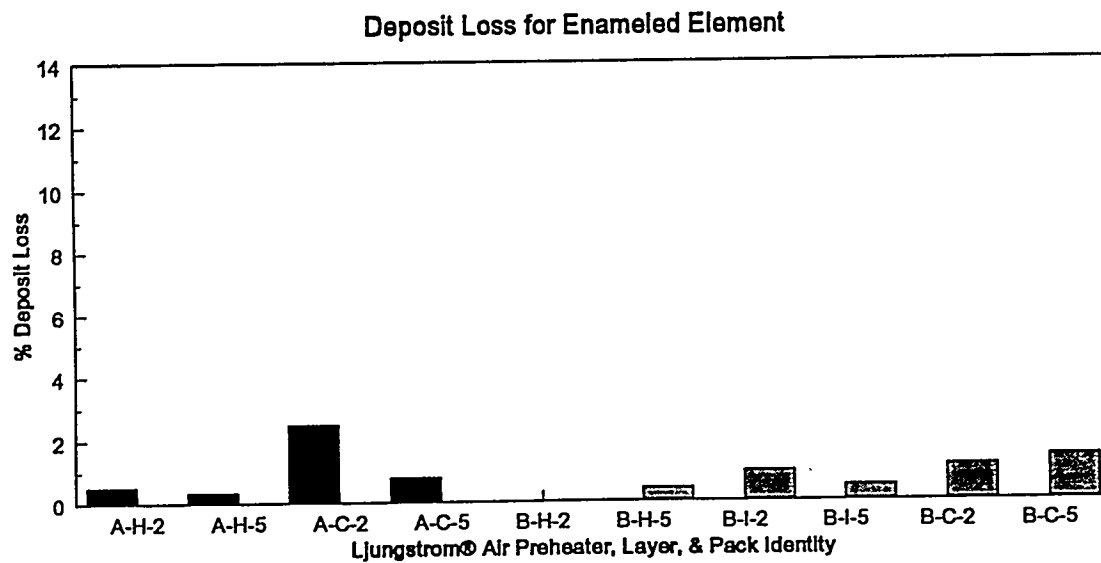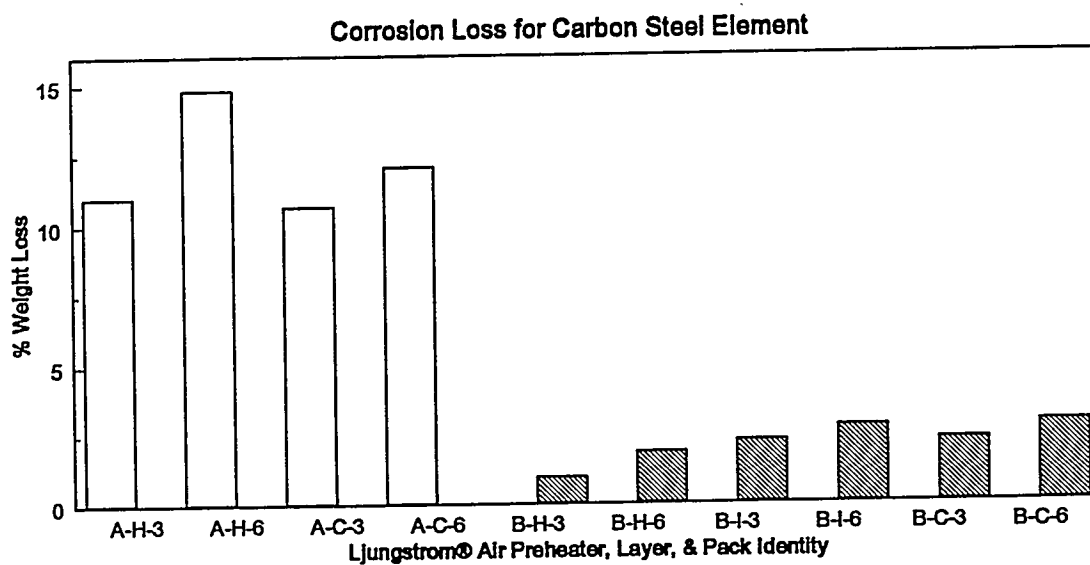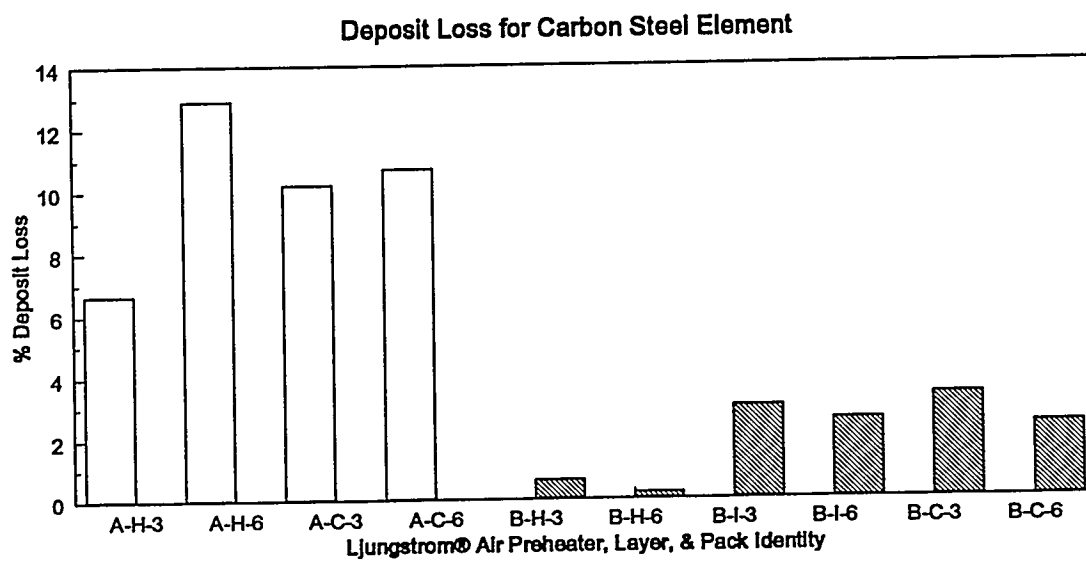
Figure 19
Metal Temperatures for Ljungstrom® Air Preheater "A"
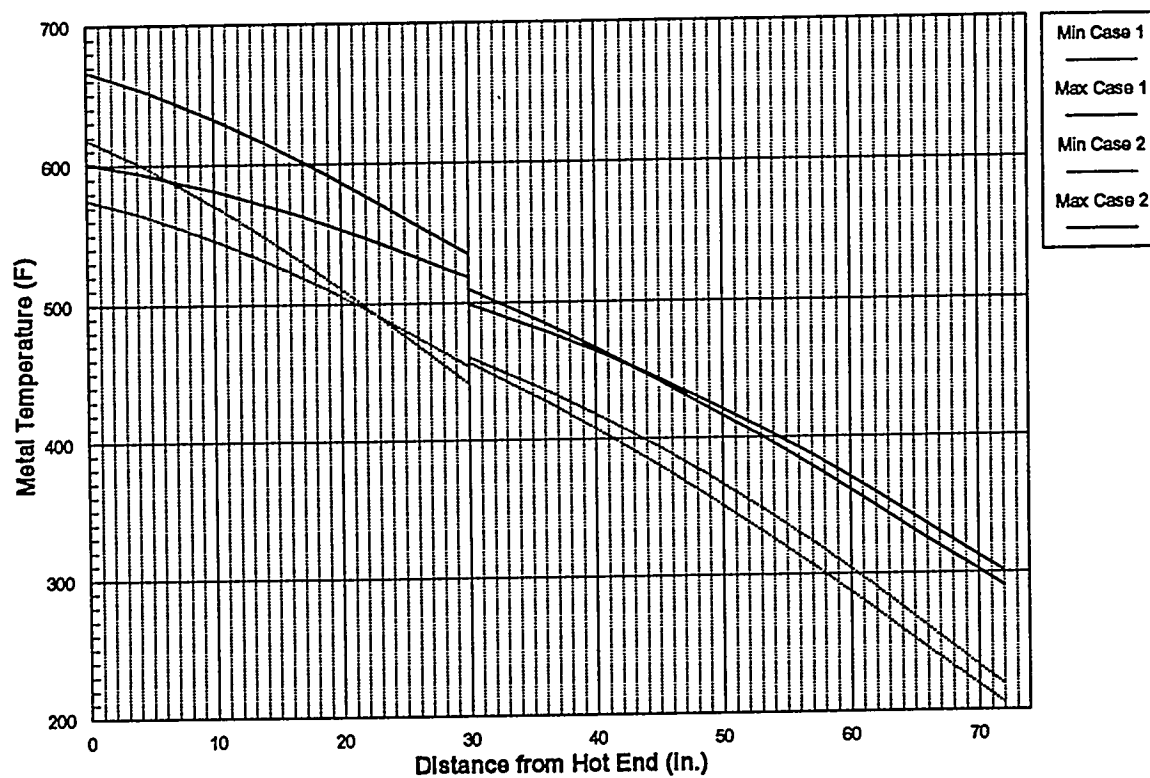
Figure 20
Metal Temperatures for Ljungstrom® Air Preheater "B"

ABB

ASEA BROWN BOVERI
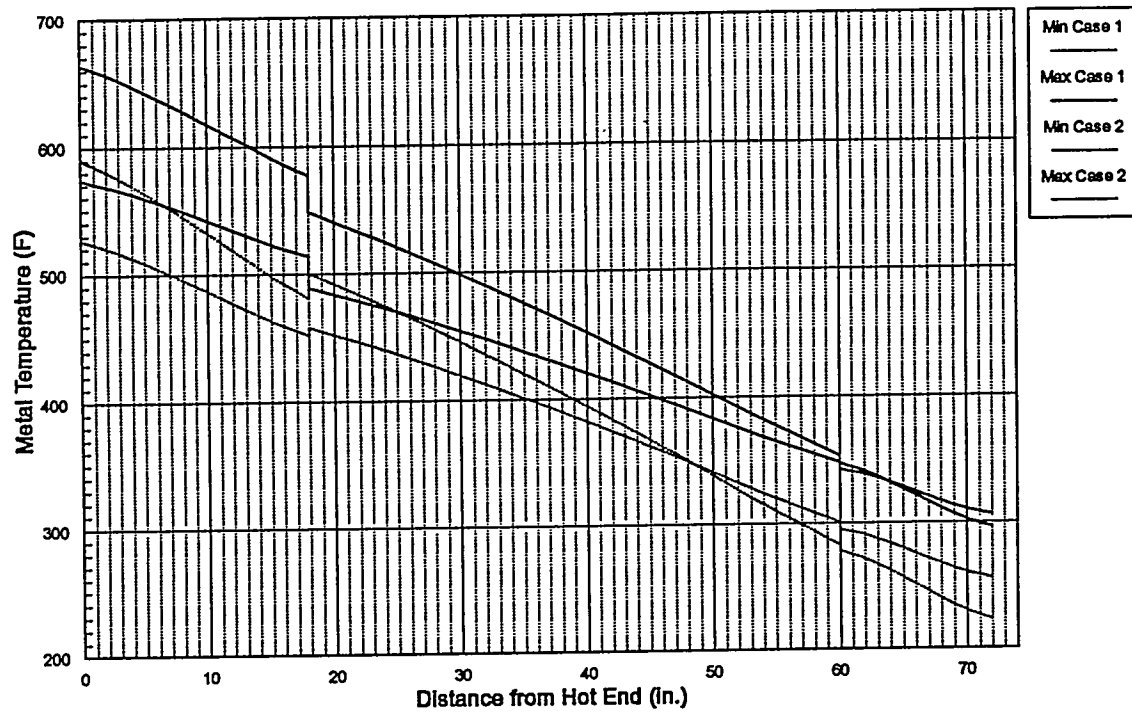
January 29, 1995

Mr. Scott Harting
Senior Matrials Engineer
ABB Air Preheater Inc.
Post Office Box 372
Wellsville, New York 14895

Subject: Final Report of the Analysis of Air Heater Deposit Samples

   Re: ABB Air Preheater
        Project 96234001 (PO P5X4023)

Dear Scott:

This is the final report of the analytical work that you in your letter of December 6, 1995. Unfortunately, we were unable to start these analyses unit after January 1 because of our laboratory relocation. I hope that this delay did not cause any serious inconvenience. Preliminary data including the x-ray diffraction (XRD) data were faxed to you on January 19, 1996.

The seven (7) samples are identified as follows:

51915-A	Sample 1 - "A" hot end top 16"	51916-A	Sample 2 "A" hot end bottom 14"
51917-A	Sample 3 "A" cold end top 5"	51918-A	Sample 4 "A" cold end top 5-14"
51919-A	Sample 5 "A" cold end 14-38"	51920-A	Sample 6 "A" cold end bottom 4"
51921-A	Sample 10 "B" intermediate bottom 20"		

These deposit samples were analyzed for ash content, carbon, hydrogen, and nitrogen content using the applicable ASTM coal standard methods. Each sample was also analyzed for XRD compound ident-ification, solution pH, and elemental composition of the ignited material. The results of these tests are enclosed in the attached data sheets and memos.

The results of the XRD analyses shows all 7 of these samples to contain hydrated iron and iron sulfate compounds. This is confirmed by their loss on ignition (ash), hydrogen content, and acidic solution pH. In addition, XRD detected the presence of ammonium iron sulfate (ammoniojarosite) in samples 51918, 51919, 51920, and 51921. Each of these samples contain nitrogen in excess of 0.1%. The gravimetric factor from nitrogen to ammoniojarosite is 34.24, meaning that each 1% nitrogen content is equivalent to 34.24% ammoniojarosite. This yields a low value of 0.7% ammoniojarosite in sample 51916-A (which is below the sensitivity of XRD) to a high of 42.8% ammoniojarosite for sample 51921-A.

The XRF elemental compoistion was measured on an ignited fraction of each sample. This data shows the samples to be predominantly iron. Their composition is consistent with ferrous corrosion products containing approximately 7% to 30% coal ash. Given the low pH of the deposit solution and the highly hydrated form of the compounds present, it is no wonder that corrosion occurs in such a wet and acidic environment.

If I can be of any furthur assistance, please call me at 203-285-5747.

Sincerely,

Kurt W. Johnson
Coordinator - PPL Chemical Analytical Services

ABB Power Plant Laboratories

To: Kurt W. Johnson/USPPL/ABB
cc:
From: Arnold L Tyler/USPPL/ABB
(Phone: + 1-203-285-2689)
Date: 01/17/96 06:16:58 PM
Subject: ABB Air Preheater, Wellsville (NY) [Ashes]

Samples of seven ashes from ABB Air Preheater, Wellsville, were submitted to the ABB-PPL X-Ray Laboratory for qualitative x-ray diffraction analyses. These samples were identified as follows:

Hot End Top 16"	PPL 5-1915-A
Hot End Bottom 14"	PPL 5-1916-A
Cold End Top 5"	PPL 5-1917-A
Cold End Top 5-14"	PPL 5-1918-A
Cold End 14-38"	PPL 5-1919-A
Cold End Bottom 4"	PPL 5-1920-A
Intermediate Bottom 20"	PPL 5-1921-A

The polycrystalline phases indicated in these analyses are given in the tables below.

ICDD#	Formula	Mineral	Confidence	Level

Hot End Top 16"   PPL 5-1915-A

ICDD#	Formula	Mineral	Confidence	Level
25-1402	Fe_2O_3	Maghemite	Definite	Major
33-0664	Fe_2O_3	Hematite	Definite	Major
25-0409	$Fe(OH)SO_4 \cdot 2H_2O$	Butlerite	Definite	Minor
31-0650	$(H_3O)Fe_3(SO_4)_2(OH)_6$	Hydronium jarosite	Definite	Minor
33-1161	SiO_2	Quartz	Possible	Trace
19-0632	$FeSO_4 \cdot 4H_2O$	Rosenite	Probable	Trace

Hot End Bottom 14"   PPL 5-1916-A

ICDD#	Formula	Mineral	Confidence	Level
39-1346	Fe_2O_3	Maghemite	Definite	Major
33-0664	Fe_2O_3	Hematite	Definite	Major

31-0650 19-0632	$(H_3O)Fe_3(SO_4)_2(OH)_6$ $FeSO_4 \cdot 4H_2O$	Hydronium jarosite Rosenite	Definite Definite	Minor Trace
25-0409	$Fe(OH)SO_4 \cdot 2H_2O$	Butlerite	Definite	Trace

Cold End Top 5"   PPL 5-1917-A

39-1346 33-0664	Fe_2O_3 Fe_2O_3	Maghemite Hematite	Definite Definite	Major Minor
31-0650 19-0632	$(H_3O)Fe_3(SO_4)_2(OH)_6$ $FeSO_4 \cdot 4H_2O$	Hydronium jarosite Rosenite	Definite Definite	Trace Trace
25-0409	$Fe(OH)SO_4 \cdot 2H_2O$	Butlerite	Definite	Trace

Cold End Top 5-14"   PPL 5-1918-A

39-1346 26-1014	Fe_2O_3 $(NH_4)Fe_3(SO_4)_2(OH)_6$	Maghemite Ammoniojarosite	Definite Definite	Major Minor
33-0664 19-0632	Fe_2O_3 $FeSO_4 \cdot 4H2O$	Hematite Rozenite	Definite Definite	Minor Trace
31-0650 29-0713	$(H_3O)Fe_3(SO_4)_2(OH)_6$ $FeO(OH)$	Hydronium jarosite Goethite	Possible Probable	Trace Trace

Cold End 14-38"   PPL 5-1919-A

39-1346 26-1014	Fe_2O_3 $(NH_4)Fe_3(SO_4)_2(OH)_6$	Maghemite Ammoniojarosite	Definite Probable	Major Minor
31-0650 29-0713	$(H_3O)Fe_3(SO_4)_2(OH)_6$ $FeO(OH)$	Hydronium jarosite Goethite	Probable Definite	Minor Trace
19-0632 25-1153	$FeSO_4 \cdot 4H_2O$ $FeFe_2(SO_4)_4 \cdot 22H_2O$	Rozenite Bilinite	Definite Possible	Trace Trace

33-1161 33-0664	SiO_2 Fe_2O_3	Quartz Hematite	Possible Possible	Trace Trace

Cold End Bottom 4"   PPL 5-1920-A

39-1346 19-0632	Fe_2O_3 $FeSO_4 \cdot 4H_2O$	Maghemite Rozenite	Definite Definite	Major Minor
33-1161 25-1153	SiO_2 $FeFe_2(SO_4)4 \cdot 22H_2O$	Quartz Bilinite	Possible Possible	Trace Trace
31-0650 26-1014	$(H_3O)Fe_3(SO_4)_2(OH)_6$ $(NH_4)Fe_3(SO_4)_2(OH)_6$	Hydronium jarosite Ammoniojarosite	Probable Possible	Trace Trace
29-0713	$FeO(OH)$	Goethite	Probable	Trace

Intermediate Bottom 20"   PPL 5-1921-A

26-1014 39-1346	$(NH_4)Fe_3(SO_4)_2(OH)_6$ Fe_2O_3	Ammoniojarosite Maghemite	Definite Probable	Major Minor
29-0713 31-0650	$FeO(OH)$ $(H_3O)Fe_3(SO_4)_2(OH)_6$	Goethite Hydronium jarosite	Definite Probable	Minor Minor
25-1153 19-0632	$FeFe_2(SO_4)_4 \cdot 22H_2O$ $FeSO_4 \cdot 4H_2O$	Bilinite Rozenite	Definite Definite	Trace Trace
33-0664 33-1161	Fe_2O_3 SiO_2	Hematite Quartz	Probable Possibe	Trace Trace

2-Theta - Scale   ABB PPL CR Analytical Services X-Ray (Diffraction) Lab 19-Jan-1996 10:13
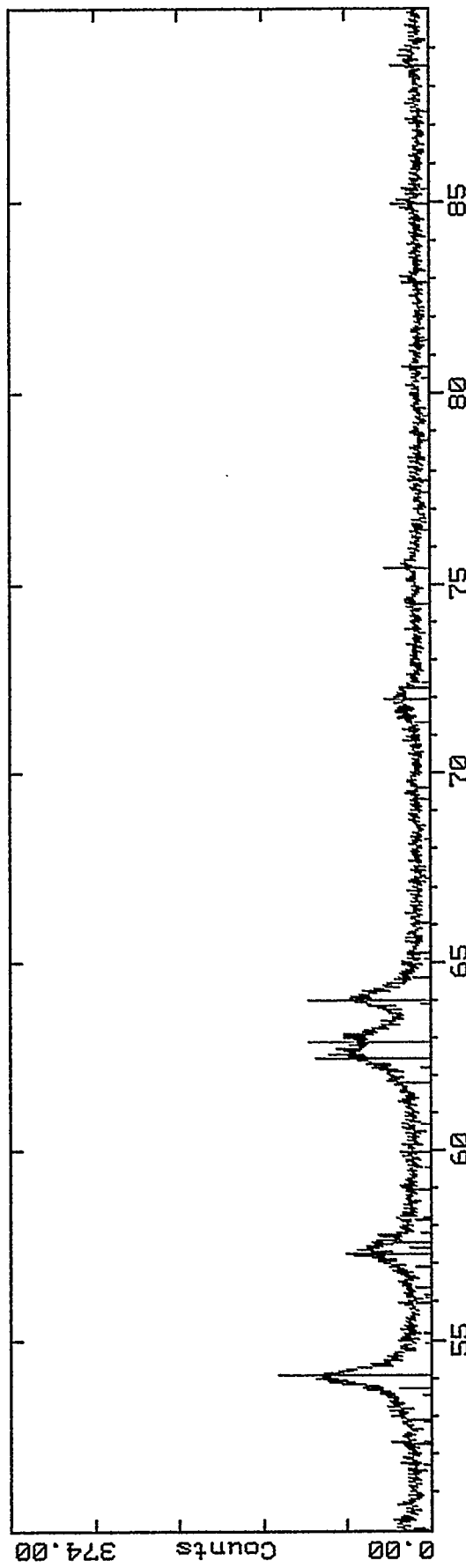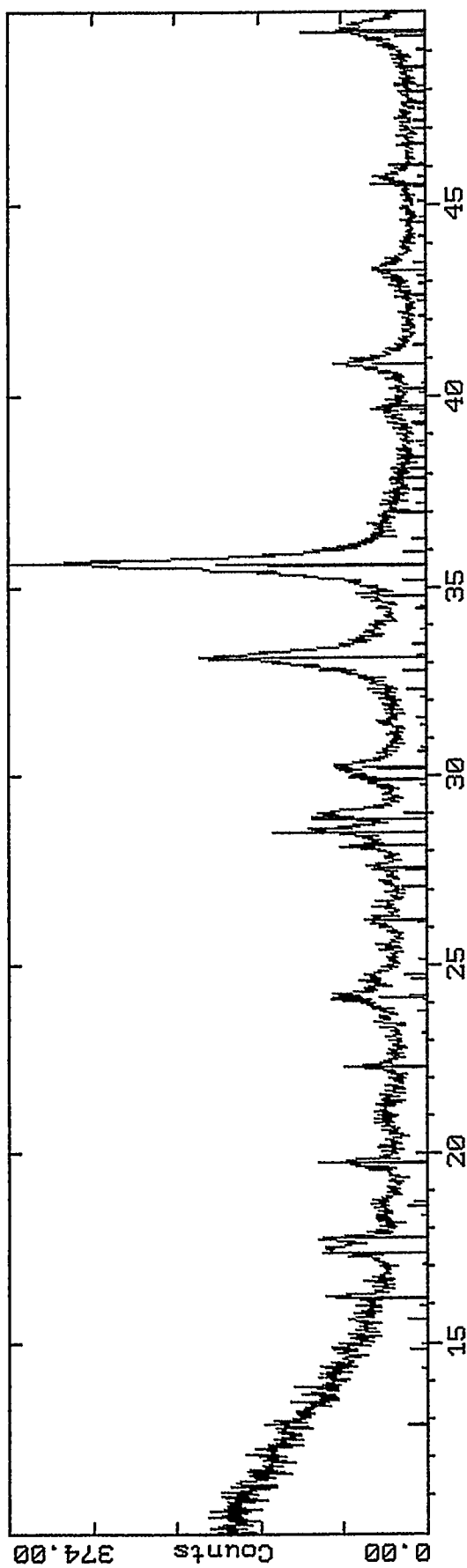
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39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
33-0664 * Fe2O3 Hematite syn (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
25-0409 * Fe(OH)SO4.2H2O Butlerite syn (WL: 1.5406Ao)

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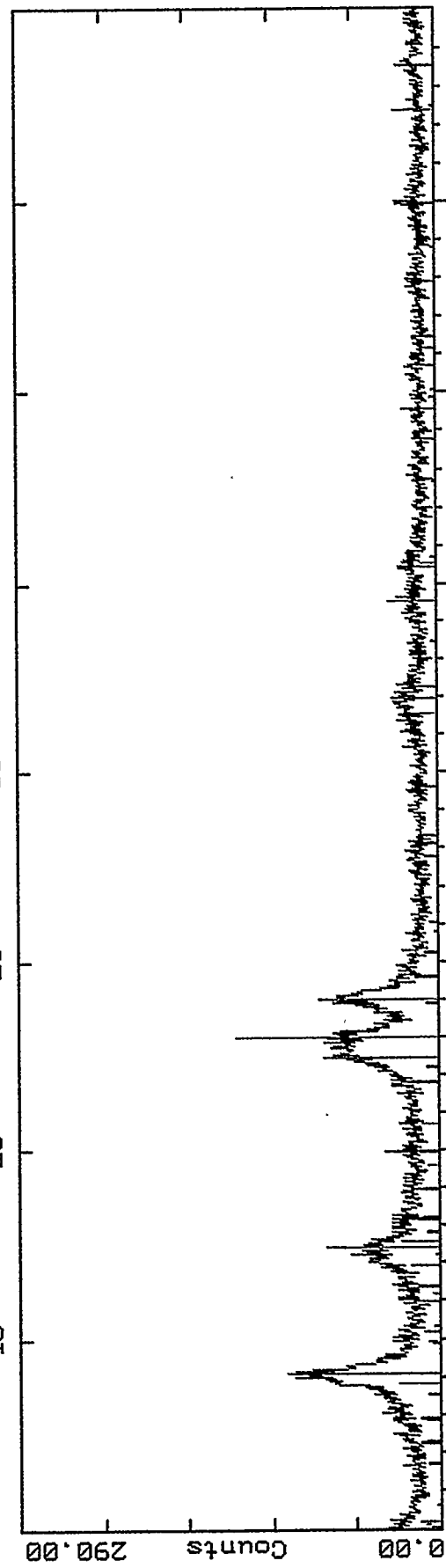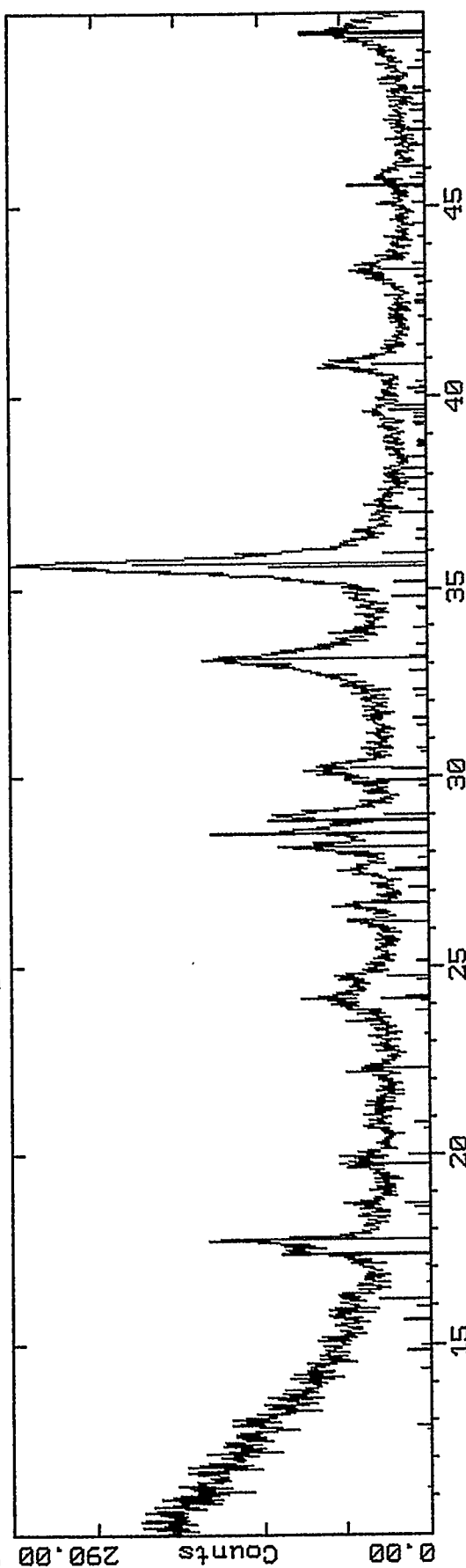
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25-1402 I Fe2O3 Maghemite ITQ RG syn (WL: 1.5406Ao)
33-0664 x Fe2O3 Hematite syn (WL: 1.5406Ao)
25-0409 x Fe(OH)SO4.2H2O Butlerite syn (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
33-1161 x SiO2 Quartz syn (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)

2-Theta - Scale   ABB PPL CR Analytical Services X-Ray (Diffraction) Lab 19-Jan-1996 10:30

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Counts   447.00   0.00
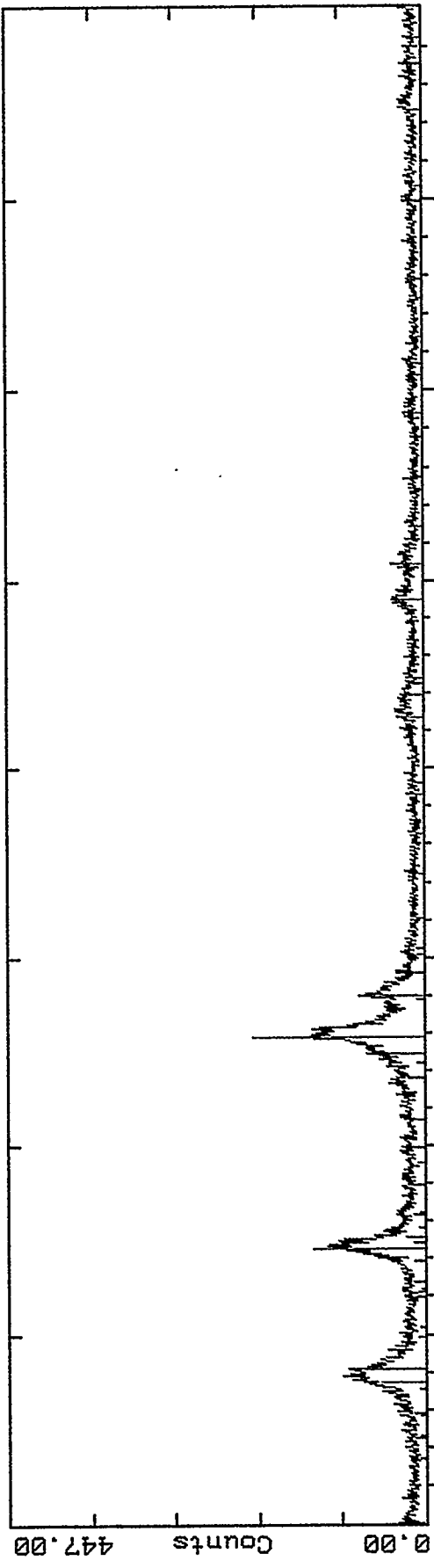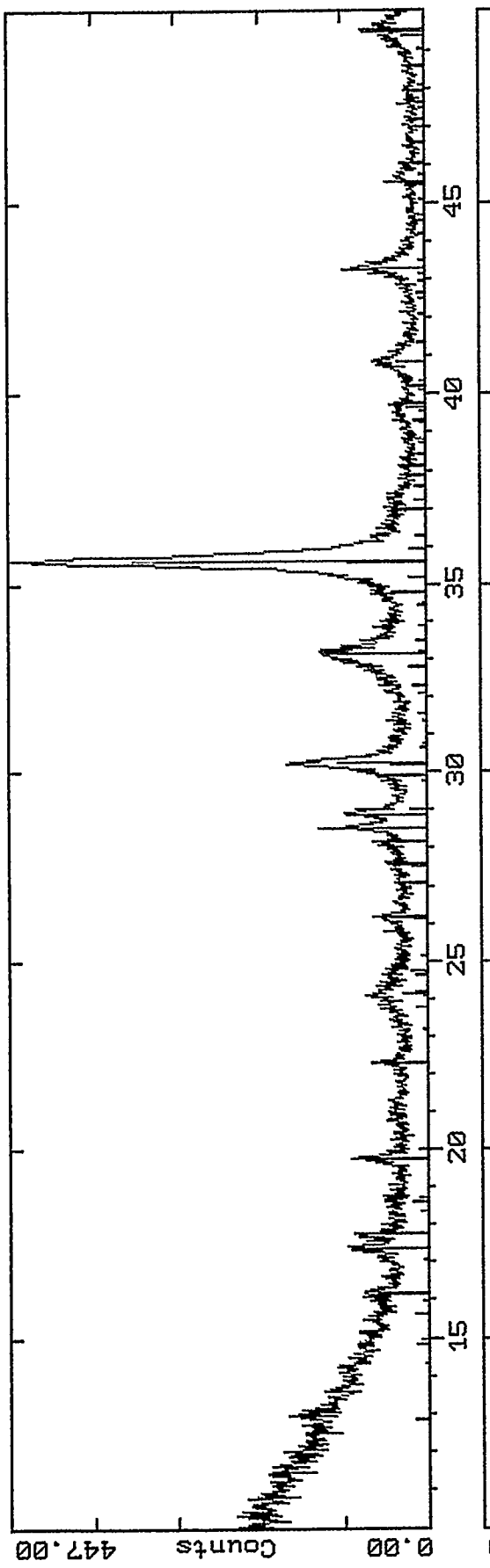
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39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
33-0664 * Fe2O3 Hematite syn (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
25-0409 * Fe(OH)SO4.2H2O Butlerite syn (WL: 1.5406Ao)

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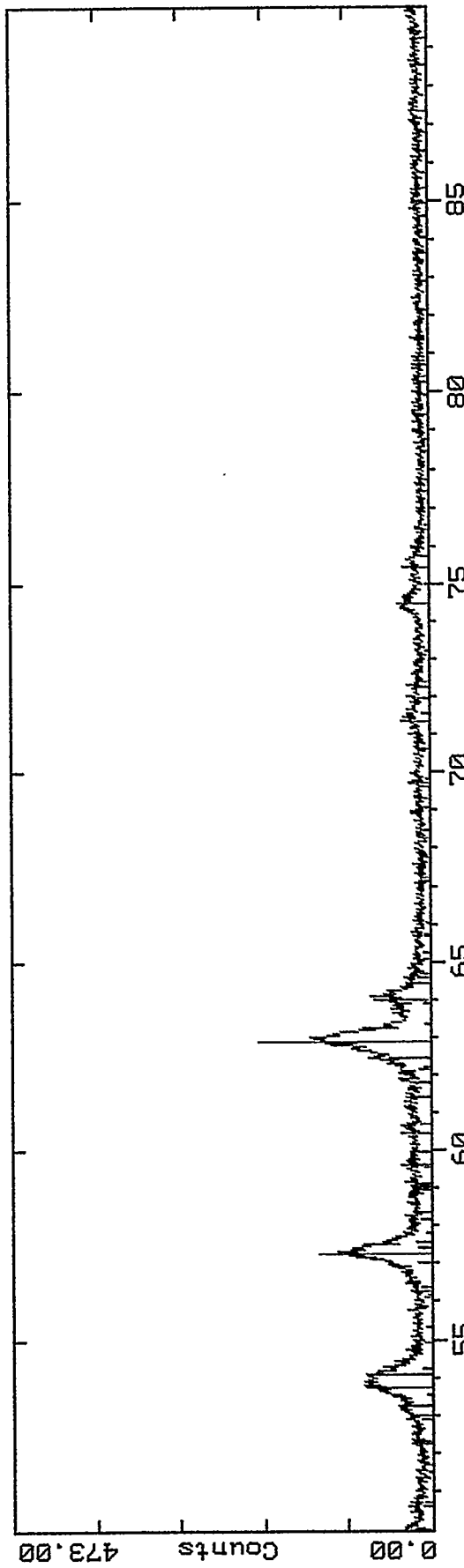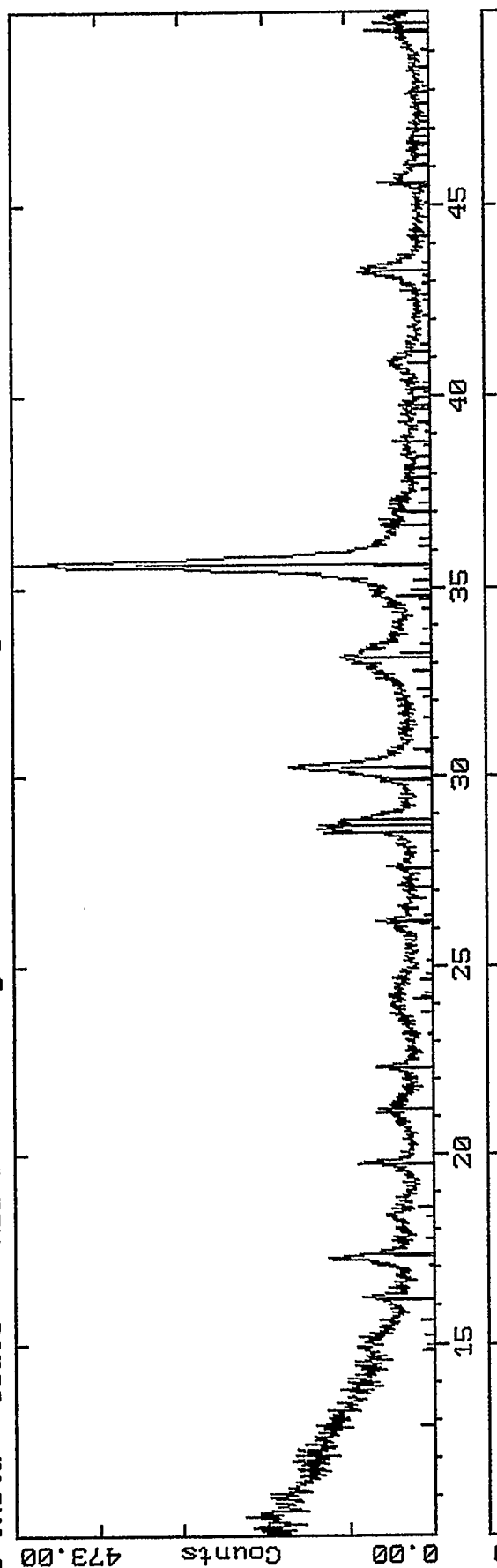
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39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
26-1014 I (NH4)Fe3(SO4)2(OH)6 Ammoniojarosite syn (WL: 1.5406Ao)
33-0664 * Fe2O3 Hematite syn (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium jarosite syn (WL: 1.5406Ao)
29-0713 I FeO(OH) Goethite (WL: 1.5406Ao)

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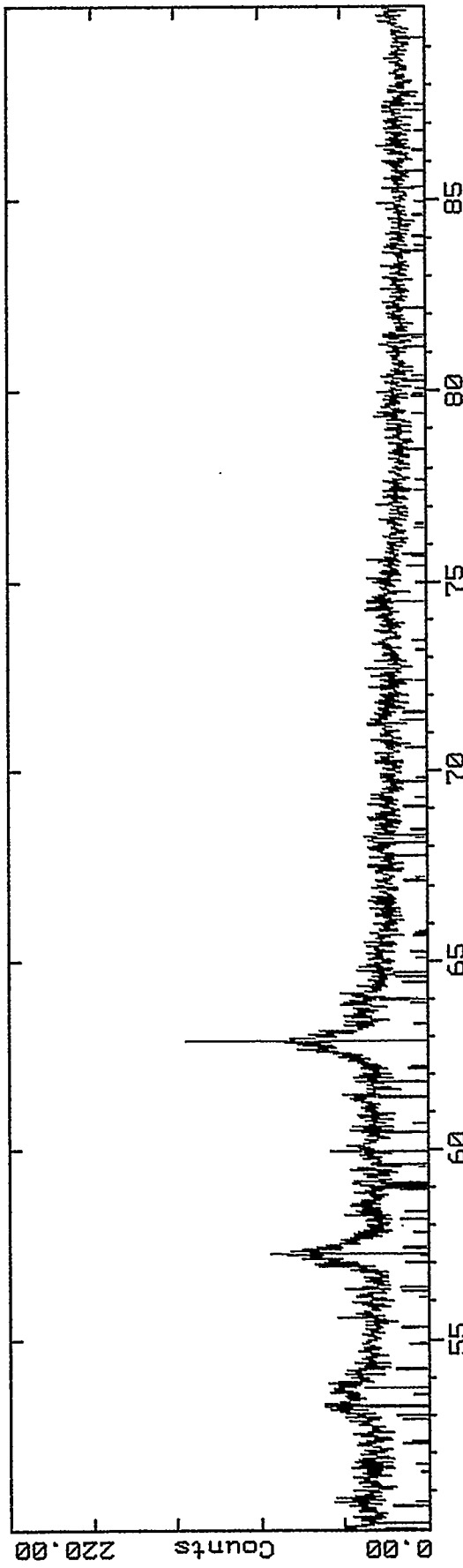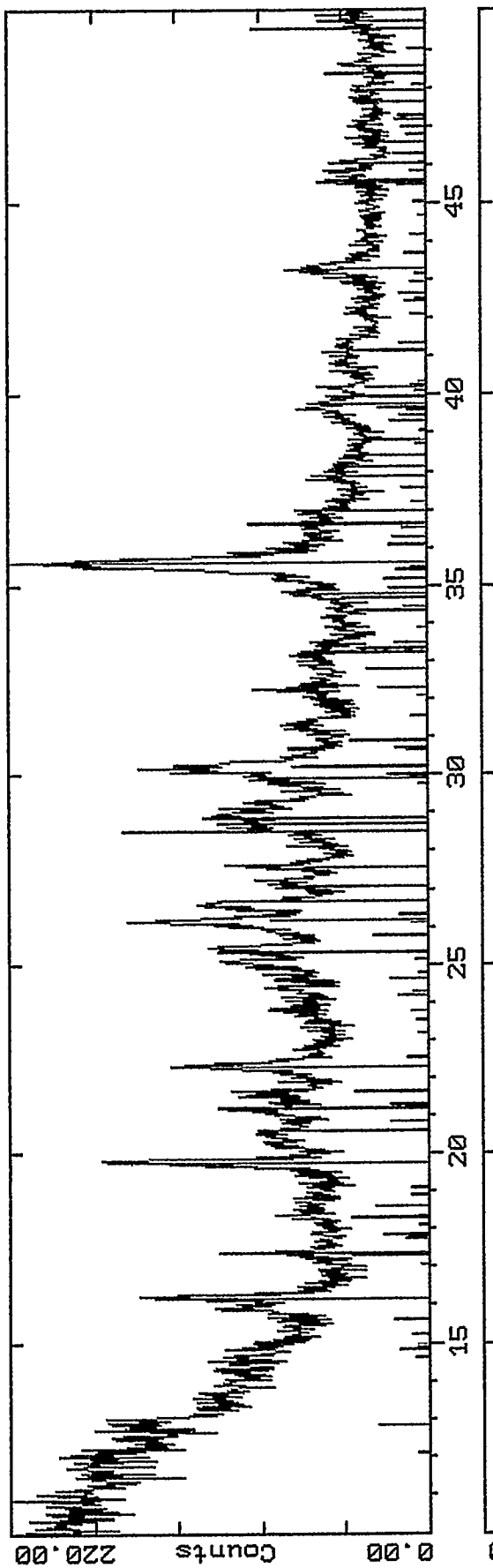
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39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
33-1161 * SiO2 Quartz syn (WL: 1.5406Ao)
25-1153 F FeFe2(SO4)4.22H2O Bilinite (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
26-1014 I (NH4)Fe3(SO4)2(OH)6 Ammoniojarosite syn (WL: 1.5406Ao)
29-0713 I FeO(OH) Goethite (WL: 1.5406Ao)

2-Theta - Scale   ABB PPL CR Analytical Services X-Ray (Diffraction) Lab 19-Jan-1996 12:08

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Counts   437.00   0.00
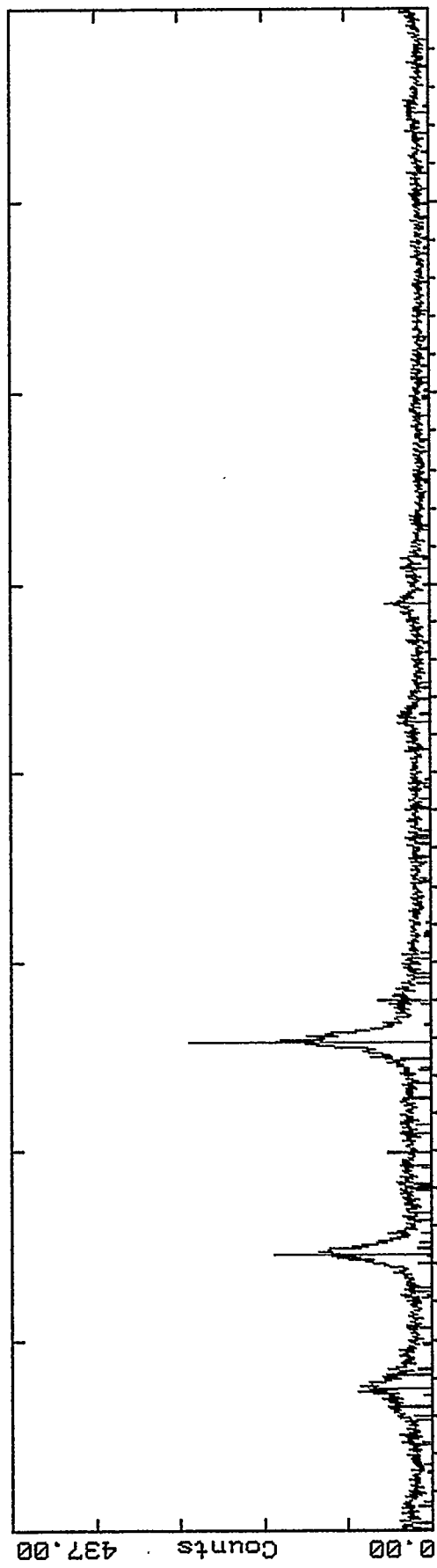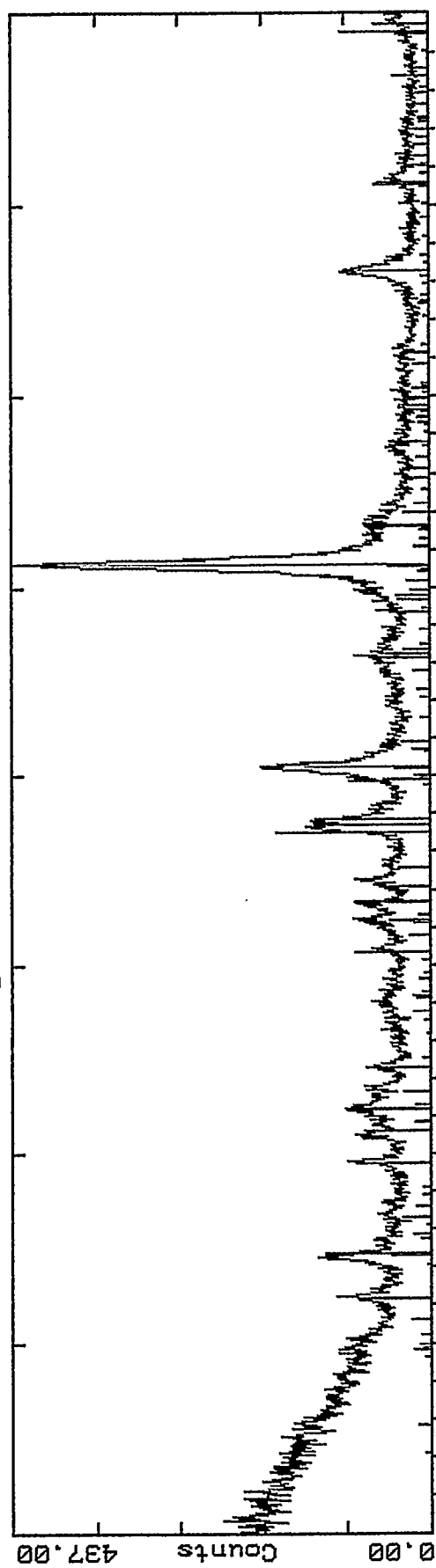
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39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
26-1014 I (NH4)Fe3(SO4)2(OH)6 Ammoniojarosite syn (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
29-0713 I FeO(OH) Goethite (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
25-1153 FeFe2(SO4)4.22H2O Bilinite (WL: 1.5406Ao)
33-1161 * SiO2 Quartz syn (WL: 1.5406Ao)
33-0664 * Fe2O3 Hematite syn (WL: 1.5406Ao)

2-Theta - Scale   ABB PPL CR Analytical Services X-Ray (Diffraction) Lab 22-Jan-1996 13:23
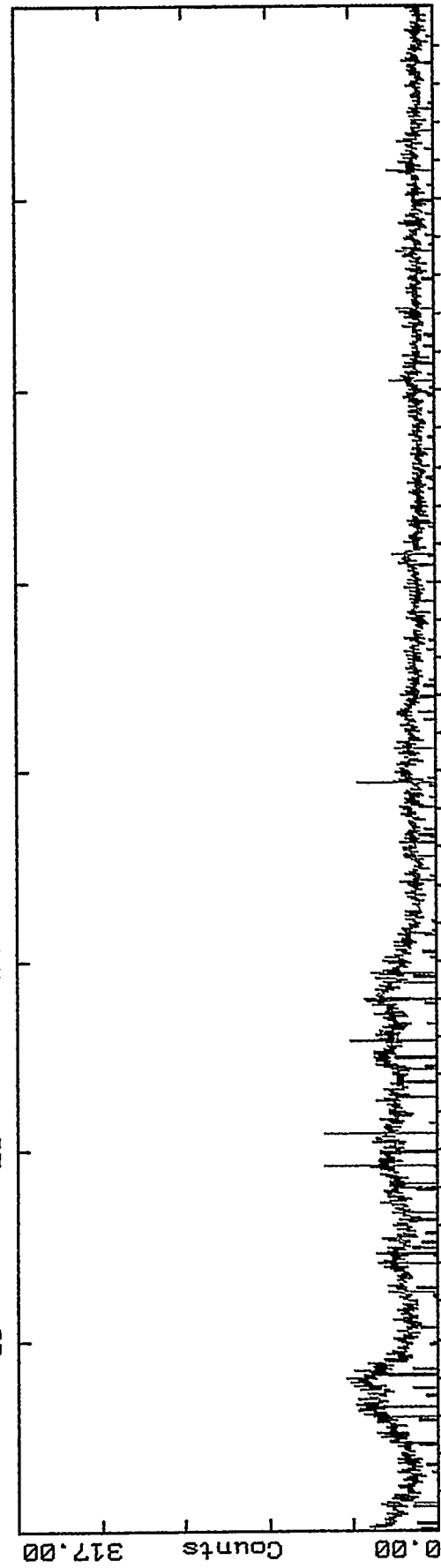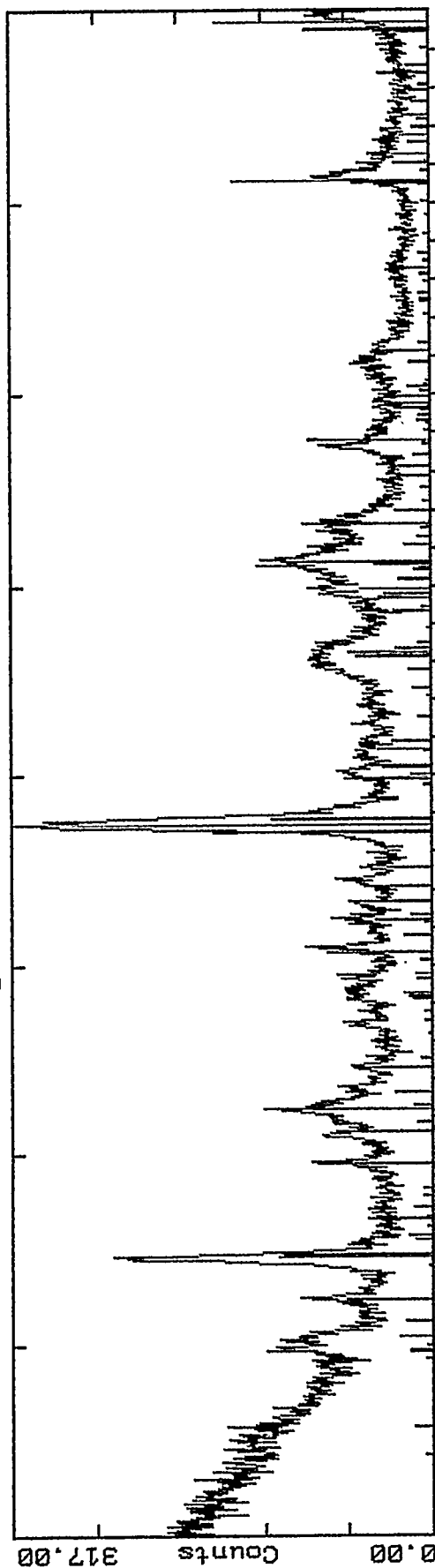
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26-1014 I (NH4)Fe3(SO4)2(OH)6 Ammoniojarosite syn (WL: 1.5406Ao)
39-1346 * Fe2O3 Maghemite ITC RG syn (WL: 1.5406Ao)
29-0713 I FeO(OH) Goethite (WL: 1.5406Ao)
31-0650 I (H3O)Fe3(SO4)2(OH)6 Hydronium Jarosite syn (WL: 1.5406Ao)
25-1153 FeFe2(SO4)4.22H2O Bilinite (WL: 1.5406Ao)
19-0632 Q FeSO4.4H2O Rozenite syn (WL: 1.5406Ao)
33-0664 * Fe2O3 Hematite syn (WL: 1.5406Ao)
33-1161 * SiO2 Quartz syn (WL: 1.5406Ao)

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

```
COMPANY   : ABB AIR PREHEATER              KDL NUMBER: 5-1915-A
LOCATION  : WELLSVILLE, NY                 ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "1" HOT END TOP 16"
CONTRACT  : P5X4023                        PROJECT   :  234001
```

```
                              AS
                              RECEIVED
                              --------

ULTIMATE ANAL., WT. PERCENT
    HYDROGEN                    1.0
    CARBON                      0.1
    NITROGEN                    0.0
    ASH                        78.6
    TOTAL                     100
```

```
            ASH COMPOSITION, WT. PERCENT
                SiO2              4.3
                Al2O3             2.2
                Fe2O3            90.4
                CaO               0.3
                MgO               0.1
                Na2O              0.4
                K2O               0.3
                TiO2              0.1
                P2O5              0.1
                SO3               0.9
                MnO               0.2
                Total            99.3
                       RATIOS
                BASE/ACID        13.86
                Fe2O3/CaO       301.33
                SiO2/Al2O3        1.95
```

```
REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 2.8
```

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER   KDL NUMBER: 5-1916-A
LOCATION : WELLSVILLE, NY   ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "2" HOT END BOTTOM 14"
CONTRACT : P5X4023   PROJECT   : 234001

	AS RECEIVED
ULTIMATE ANAL., WT. PERCENT	
HYDROGEN	1.0
CARBON	0.0
NITROGEN	0.0
ASH	80.4
TOTAL	100

ASH COMPOSITION, WT. PERCENT	
SiO_2	2.5
Al_2O_3	1.4
Fe_2O_3	94.3
CaO	0.1
MgO	0.1
Na_2O	0.3
K_2O	0.2
TiO_2	0.1
P_2O_5	< 0.1
SO_3	0.6
MnO	0.2
Total	99.8
RATIOS	
BASE/ACID	23.75
Fe_2O_3/CaO	943.0
SiO_2/Al_2O_3	1.79

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 3.0

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER   KDL NUMBER: 5-1917-A
LOCATION : WELLSVILLE, NY   ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "3" COLD END TOP 5"
CONTRACT : P5X4023   PROJECT   : 234001

```
                              AS
                            RECEIVED
                            --------
ULTIMATE ANAL., WT. PERCENT
   HYDROGEN                    0.7
   CARBON                      0.0
   NITROGEN           .        0.1
   ASH                        83.6
   TOTAL                      100


         ASH COMPOSITION, WT. PERCENT
            SiO2                1.7
            Al2O3               1.1
            Fe2O3              95.7
            CaO              < 0.1
            MgO                 0.1
            Na2O                0.3
            K2O                 0.1
            TiO2             < 0.1
            P2O5             < 0.1
            SO3                 0.6
            MnO                 0.2
            Total              99.9
                     RATIOS
            BASE/ACID          33.89
            Fe2O3/CaO        2392.50
            SiO2/Al2O3          1.55
```

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 3.2

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER   KDL NUMBER: 5-1918-A
LOCATION : WELLSVILLE, NY   ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "4" COLD END TOP 5-14"
CONTRACT : P5X4023   PROJECT   : 234001

	AS RECEIVED
ULTIMATE ANAL., WT. PERCENT	
HYDROGEN	0.8
CARBON	0.1
NITROGEN	0.3
ASH	84.7
TOTAL	100

ASH COMPOSITION, WT. PERCENT	
SiO_2	0.9
Al_2O_3	0.8
Fe_2O_3	96.9
CaO	< 0.1
MgO	0.1
Na_2O	0.2
K_2O	0.1
TiO_2	0.1
P_2O_5	< 0.1
SO_3	0.6
MnO	0.2
Total	100.0
RATIOS	
BASE/ACID	54.08
Fe_2O_3/CaO	2422.50
SiO_2/Al_2O_3	1.13

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 3.4

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER
LOCATION : WELLSVILLE, NY
SAMPLE ID: SAMPLE "5" COLD END 14-38"
CONTRACT : P5X4023

KDL NUMBER: 5-1919-A
ANAL DATE : 01/29/96

PROJECT   : 234001

	AS RECEIVED
ULTIMATE ANAL., WT. PERCENT	
HYDROGEN	1.4
CARBON	0.3
NITROGEN	0.8
ASH	75.8
TOTAL	100

ASH COMPOSITION, WT. PERCENT	
SiO2	6.8
Al2O3	4.1
Fe2O3	85.1
CaO	0.6
MgO	0.3
Na2O	0.4
K2O	0.5
TiO2	0.2
P2O5	0.1
SO3	1.3
MnO	0.2
Total	99.6

RATIOS	
BASE/ACID	7.83
Fe2O3/CaO	141.83
SiO2/Al2O3	1.66

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 3.2

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER   KDL NUMBER: 5-1920-A
LOCATION : WELLSVILLE, NY   ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "6" COLD END BOTTOM 4"
CONTRACT : P5X4023   PROJECT   : 234001

	AS RECEIVED
ULTIMATE ANAL., WT. PERCENT	
HYDROGEN	2.2
CARBON	0.5
NITROGEN	0.4
ASH	55.5
TOTAL	100

ASH COMPOSITION, WT. PERCENT	
SiO_2	15.3
Al_2O_3	8.4
Fe_2O_3	68.1
CaO	1.1
MgO	0.7
Na_2O	1.2
K_2O	1.1
TiO_2	0.5
P_2O_5	0.3
SO_3	2.2
MnO	0.1
Total	99.0
RATIOS	
BASE/ACID	2.98
Fe_2O_3/CaO	61.91
SiO_2/Al_2O_3	1.82

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 2.8

ABB COMBUSTION ENGINEERING
POWER PLANT LABORATORIES CHEMICAL ANALYTICAL SERVICES
WINDSOR, CONNECTICUT
FUEL ANALYSIS REPORT

COMPANY : ABB AIR PREHEATER   KDL NUMBER: 5-1921-A
LOCATION : WELLSVILLE, NY   ANAL DATE : 01/29/96
SAMPLE ID: SAMPLE "10" B INTERMEDIATE BOTTOM 4"
CONTRACT : P5X4023   PROJECT   : 234001

	AS RECEIVED
ULTIMATE ANAL., WT. PERCENT	
HYDROGEN	1.9
CARBON	0.4
NITROGEN	1.3
ASH	66.4
TOTAL	100

ASH COMPOSITION, WT. PERCENT	
SiO2	8.8
Al2O3	5.4
Fe2O3	79.5
CaO	1.3
MgO	0.4
Na2O	0.5
K2O	0.8
TiO2	0.4
P2O5	0.2
SO3	2.2
MnO	0.2
Total	99.7
RATIOS	
BASE/ACID	5.65
Fe2O3/CaO	61.15
SiO2/Al2O3	1.63

REMARKS/OTHER DATA
pH OF A 1% SOLUTION = 3.5

TESTING AND ANALYTICAL SERVICES FOR THE INNOVATIVE CLEAN COAL TECHNOLOGY DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION (SCR) TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO$_x$) EMISSIONS FROM HIGH SULFUR COAL

Final Report

Effects of SCR Ammonia on Ammonia Volatilization, Ammonia Extraction, and Metals Extraction from SCR Fly Ash

Prepared by

Kenneth M. Cushing
Robert F. Heaphy
Edward B. Dismukes

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Prepared for

SOUTHERN COMPANY SERVICES, INC.
P.O. Box 2625
Birmingham, Alabama 35202-2625

DOE Prime Contract Number DE-FC22-89-PC89652
SCS Contract Number C-92-000320
SRI Project Number 7613
SRI Report Number SRI-ENV-95-415-7613

October 20, 1995

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

ACKNOWLEDGEMENTS

Section 1

EXECUTIVE SUMMARY

This report summarizes the results of a laboratory study, conducted at Southern Research Institute for Southern Company Services, Inc., to evaluate the effects of SCR ammonia on ammonia volatilization, ammonia extraction, and metals extractability from fly ash. To conduct the study, samples of pre-reactor (ammonia-free) and post-reactor (ammonia-exposed) fly ash were collected at SCR Reactors B and C at Plant Crist.

Ammonia Extractability and Volatilization from SCR Fly Ash

Almost no ammonia volatilized from the SCR ash until a significant amount of water vapor was absorbed by the ash. A plausible mechanism for the apparent volatilization that occurred is that enough water was gained by the ash to form a moist layer with a pH high enough to evolve gas-phase ammonia from the ammonium compounds on the ash. Nearly all of the ammonia on the ash evolved to the gas phase in the closed-container experiments. Ammonia concentrations in enclosed spaces depend on the ammonia concentration of the ash, the volume of air above the ash available for dilution, and the presence of a humid atmosphere.

The extraction of ammonia from fly ash seems to depend upon pH. Evidently, all or nearly all of the ammonia present was extracted in the buffered solutions at pH 4.7 and pH 6.2, but not all was recovered in alkaline unbuffered extracts. In the pH 6.2 buffer, however, the completeness of extraction seemed to fall off somewhat as the ratio of ash to buffer increased. At 3 g of ash per 50 mL of pH 6.2 buffer, the amount of ammonia extracted was about 200 µg/g, whereas at 1 g per 50 mL, the amount was near 250 µg/g.

Ammonia concentration in the ash was much higher for the smaller particle sizes, but most of the total ammonia was found to reside with the larger particles simply because these comprise the vast majority of the ash mass. The implication is that very little slip ammonia will exit the process when high efficiency particulate emission controls are in place since all detectable ammonia is in the solid phase at the air heater exit and most of the ammonia is associated with the larger particle sizes which are most readily collected.

Metals Extractability from SCR Fly Ash

The SCR fly ash samples were subjected to extraction with water, and the extracts were analyzed for each of 28 metals. This was done to ascertain whether exposure of the fly

ash to ammonia vapor caused an enhancement of the metals extractabilities under conditions resembling those that might exist in an ash pond.

Of the 28 metals included in the study, only 17 could be detected in the fly-ash extracts. Of these 17 detectable metals, only barium underwent an increase in extractability following exposure to ammonia. The magnitude of the increase was found to depend directly on the magnitude of the NH_3/NO_x ratio in the SCR unit, however, the increase was slight for all NH_3/NO_x ratios tested. Of the 16 additional metals that could be detected in the fly-ash extracts, none displayed what we considered to be genuine enhancements in extractability, and several exhibited decreases in extractability as a result of exposure of the fly ash to ammonia. Although one of these metals -- Se -- displayed a large apparent increase in extractability on exposure to ammonia, we concluded that the selenium found in the reactor-outlet sample extracts must have condensed from the gas phase onto the fly ash at the reactor outlet. Finally, a deliberate downward adjustment in the pH of one sample solution caused enhancements in the extractabilities of several metals, most notably Mg, but also Mn, Ca, As and Fe to a lesser degree.

Section 2

INTRODUCTION

The Selective Catalytic Reduction (SCR) process is a pollution-control strategy that is designed to reduce the emissions of NO_x from power plants. In this process, gaseous ammonia is injected into the flue-gas stream, where it chemically reduces the NO_x to nitrogen gas at an elevated temperature. The SCR process has recently been evaluated at DOE's SCR Demonstration Plant at Gulf Power Company's Plant Crist.

Unfortunately, the ammonia added in the SCR process does not completely react with NO_x. Of course, it is possible to increase the efficiency of NO_x removal by increasing the amount of ammonia (i.e., by boosting the NH_3/NO_x ratio in the reactor). But this also increases the amount of excess, unreacted ammonia in the flue stream. Much of this excess ammonia is known to be taken up by the entrained fly ash, however, this almost always occurs because of a reduction in flue gas temperature as happens in an air heater. The injection of ammonia in the presence of fly ash at 700°F may not cause any ammonia to deposit or be absorbed on the fly ash. Nevertheless, the effects of absorbed ammonia on the properties of the fly ash are not well understood.

One of the major concerns about fly ash that has been exposed to ammonia in the SCR process is how the ammonia in the ash may be extracted and what will be the resulting ammonia concentration measured in the laboratory. The species is likely to occur in the ash as ammonium sulfate $(NH_4)_2SO_4$. Either the ammonium ion (NH_4^+) or free ammonia (NH_3) is likely to be easily dissolved in water. The completeness of extraction, however, may depend on the pH of the extracting medium, which regulates the balance between NH_4^+ and NH_3 in the extract. Either the innate alkalinity of the ash and the pH the ash produces when wet, or the pH of a buffer used for extraction, may influence the efficiency of ammonia removal from the ash.

Another concern is the volatility of ammonia in ash. In the form NH_4^+ there is likely to be only an infinitesimal volatility. Moisture in air, however, is capable of hydrolyzing the ammonium ion and forming the much more volatile NH_3 species:

$$NH_4^+ \text{ (aq)} + H_2O \text{ (l)} = NH_3 \text{ (aq)} + H_3O^+ \text{ (aq)}$$
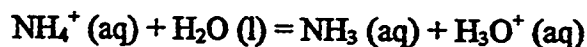
The hydronium ion, or hydrated proton shown as a product in this reaction, will be neutralized by alkalinity in the ash, and thus the reaction made to proceed further toward completion. Air with a high relative humidity which may lead to a high concentration of adsorbed moisture, or liquid water that actually wets the ash, are likely to increase volatilization.

Still another major concern about the presence of ammonia in the fly ash is whether the ammonia-influences the water extractability of pollutant metals from the ash. That is, will the presence of ammonia in the ash lead to an increase in the amount of dissolved metals in the ash ponds? Clearly, the answer to this question could have a significant impact on the ability of a power plant to meet the requirements of its regulatory pollutant-discharge permits.

This report describes a brief laboratory study performed at Southern Research Institute (Southern Research), for Southern Company Services, Inc., to address the questions of ammonia extractability, ammonia volatilization, and metals extractability from ammonia-exposed fly ash. Samples of fly ash were collected both upstream and downstream from SCR Reactor B and C at Plant Crist. Because the upstream ash samples had not been exposed to ammonia, their presence in this study allowed a direct comparison between ammonia-free and ammonia-containing ash from the same flue gas stream, as well as the attendant effects of temperature reduction as the flue gas passed through the air heater.

Section 3 of this report describes the various ash samples collected for these studies. Section 4 describes the results of the ammonia extractability and volatilization tests. The results of tests on metals extractability from SCR fly ash are summarized in Section 5. Test data are presented in various tables throughout each section.

Section 3

DESCRIPTION OF ASH SAMPLES

Ash samples for this work were obtained during the last two weeks of January 1995 during routine operation of the Plant Crist SCR Demonstration Plant. Three bulk samples of hopper ash were collected from the cyclones located downstream from the air heaters on large SCR reactors B and C. These ash samples were collected from the cyclone hoppers following a full day of reactor operation at one of three SCR parametric operating conditions. A full day of operation prior to sample collection at each test condition was required to provide sufficient time for ammonia equilibrium to be achieved. The reactor's flow rate and temperature were identical at all three parametric conditions - 5000 scfm (wet) and 700 °F, respectively. Only the ammonia to nitrogen oxides ratio (NH_3/NO_x) differed. The NH_3/NO_x ratios were 0.6 (Test Condition 21), 0.8 (Test Condition 22), or 1.0 (Test Condition 24). Hopper samples were collected from the Reactor B cyclone hopper for Test Conditions 21 and 24. For Test Condition 22 a cyclone hopper sample was collected from Reactor C.

A second set of fourteen ash samples was collected from Reactor B while operating at Test Conditions 21, 22, and 24. Table 3-1 provides information about each sample (test location, date of test, start time, end time, etc.). The samples were collected isokinetically using an in-stack filter simultaneously at two locations. Fly ash samples upstream of ammonia injection were collected at test ports located downstream of the reactor venturi flow meter. Ash samples downstream of ammonia injection were collected at the outlet of the Reactor B air heater.

A third set of size-segregated ash samples was collected at the outlet of the Reactor B air heater using a Southern Research five-stage cyclone, an in-stack sampling device. This device separated the particulate matter into five size fractions as it was collected. Since the cross-section of the duct at the test point was quite small (1 ft by 3 ft) in relation to the sampling device which is more than two feet in length, the samples were collected isokinetically at the center point of the duct. Two runs were completed while the reactor operated at parametric test condition 22, the normal baseline operating condition (a flue gas flow rate of 5000 scfm (wet), a flue gas temperature of 700 °F, a NH_3/NO_x ratio of 0.8).

Table 3-1. Identification of Reactor B inlet and air heater outlet ash samples.

Sample Code Number*	Test Identification Number*	Test Date	Start Time	End Time	NH_3/NO_x
BI 2	108-BI-M17-02	1/23/95	1530	1630	na
BI 3	108-BI-M17-03	1/24/95	1020	1220	na
BI 4	108-BI-M17-04	1/24/95	1445	1630	na
BI 5	108-BI-M17-05	1/25/95	930	1130	na
BI 6	108-BI-M17-06	1/25/95	1500	1630	na
BI 7	108-BI-M17-07	1/26/95	1010	1210	na
BI 8	108-BI-M17-08	1/26/95	1430	1615	na
BAO 2	108-BAO-M17-02	1/23/95	1535	1635	0.8
BAO 3	108-BAO-M17-03	1/24/95	1020	1220	0.8
BAO 4	108-BAO-M17-04	1/24/95	1446	1617	0.8
BAO 5	108-BAO-M17-05	1/25/95	930	1130	0.6
BAO 6	108-BAO-M17-06	1/25/95	1500	1630	0.6
BAO 7	108-BAO-M17-07	1/26/95	1010	1210	1.0
BAO 8	108-BAO-M17-08	1/26/95	1430	1615	1.0
Cyclone 1	128-BAO-CYC-01	6/17/95	959	1059	0.8
Cyclone 2	128-BAO-CYC-02	6/17/95	1330	1430	0.8

*— BI - Reactor B Inlet; BAO - Reactor B Air Heater Outlet
na — Not applicable to inlet tests.

Section 4

INVESTIGATION OF AMMONIA EXTRACTABILITY AND
VOLATILIZATION FROM FLY ASH

Prior to the volatilization and extraction studies, the ammonia content of the hopper ash samples was determined by extracting a one-gram sample of the ash in fifty milliliters of deionized water to which 4 drops of 1:1 sulfuric acid had been added. Ammonia concentrations were determined by the ion-specific-electrode method. The results of these extractions are shown in Table 4-1. As expected, higher ammonia concentrations were measured for the fly ash samples collected during operation at higher NH_3/NO_x ratios. (It is important to note that the addition of 1:1 sulfuric acid, as described here, makes a pH of about 1.7 in the ash slurry. This pH is substantially more acidic than that in either of the buffers described later for controlling the pH during ammonia extraction.)

An ammonia train was run at the Reactor B air heater outlet after the collection of the seven isokinetic ash samples to measure total ammonia in the gas stream. All ammonia concentrations were determined by the ion-specific-electrode method. The total ammonia concentration measured by the ammonia train and the solid phase ammonia concentration from the ash sample collected at the outlet of the air heater were used to determine ammonia partitioning between the gas and solid phases. Table 4-2 shows the results of these analyses and the resulting ammonia partitioning. Solid-phase ammonia concentrations are shown both on a mass fraction basis and a volume basis corrected to 3% O_2 (dry). The gas-phase ammonia concentration was calculated as the difference between the total ammonia and solid-phase ammonia concentrations. Please note that the samples were collected on different days. This could account for some of the variability observed in the ammonia partitioning. The average ratio of solid-phase ammonia to gas-phase ammonia was 0.53, 3.3, and 0.38 for NH_3/NO_x ratios of 0.6, 0.8, and 1.0, respectively.

Solid-Phase Ammonia Concentration and Ash Slurry pH

Two preliminary analyses of the isokinetic ash samples were performed before they were forwarded to Southern Research's Birmingham laboratories for trace metals analysis. The pH of each sample when 0.1 g of ash from the filter catch was mixed with 50 mL of deionized water was measured. The mixture of ash and water was placed in a sealed bottle and agitated for four hours before measuring the pH. The pH of the ash-water mixtures for the samples collected upstream of ammonia injection ranged from 7.90 to 9.94. For the ash samples collected at the air heater outlet for NH_3/NO_x ratios of 0.6, 0.8, and 1.0, the average pH values were 9.20, 9.34, and 9.77, respectively. Also, one-gram samples of the ash were placed in 50 mL of deionized water with four drops of 1:1 sulfuric

Table 4-1. Ammonia concentrations of the SCR hopper ash samples used in the ammonia volatilization experiments.

Parametric Condition	SCR Reactor	NH_3/NO_x	Reactor Flow Rate, scfm, wet	Reactor Inlet Temperature, °F	Average NH_3 * Conc., µg/g
21	B	0.6	5,000	700	48
22	C	0.8	5,000	700	254
24	B	1.0	5,000	700	352

*: Extractions performed by placing 1 g of ash in a beaker, adding 50.0 ml D.I. H_2O, 4 drops of 1:1 H_2SO_4, and stirring for 5 minutes.

Table 4-2. Partitioning of ammonia slip between gas phase and solid phase at the outlet of the B reactor air heater.

Sample Number Code	NH_3/NO_x	Total NH_3, ppm(v) @ $3\%O_2$ (dry)	Mass Conc., gr/dscf @ $3\%O_2$ (dry)	Solid-Phase NH_3, µg/g	Solid-Phase NH_3, ppm(v) @ $3\%O_2$ (dry)	Gas-Phase NH_3, ppm(v) @ $3\%O_2$ (dry)
BAO 5	0.8	2.9	2.74	93	1.0	1.9
BAO 6	0.6	2.9	3.13	79	1.0	1.9
BAO 2	0.8	3.5	3.10	185	2.5	1.0
BAO 3	0.8	3.3	2.89	238	2.9	0.3
BAO 4	0.8	3.2	3.14	175	2.2	1.0
BAO 7	1.0	32.1	2.83	888	10.2	21.9
BAO 8	1.0	34.6	2.68	735	8.0	26.6

acid, stirred for 5 minutes, and the ammonia concentration measured with an ion-specific electrode (these data are also included in Table 4-2). As expected, no ammonia (generally less than the detection limit) was found in the ash samples collected upstream of ammonia injection. Average ammonia concentrations in the ash samples collected at the air heater outlet for NH_3/NO_x ratios of 0.6, 0.8, and 1.0 were 86, 199, and 812 µg/g, respectively. The results of these analyses are shown in Table 4-3.

Also included in this table are the measured pH and ammonia concentrations for the three cyclone hopper samples from Reactors B and C. The pH values for the hopper ash samples were slightly higher on average (10.23) than those from the air heater outlet (9.42). A possible explanation for this is that the cyclone hopper ash is coarser in particle size, has a lower surface/mass ratio, contains less adsorbed SO_3, and therefore is more alkaline. The ammonia concentrations shown in Table 4-3 are different from those in Table 4-1 because separate analyses were performed. For comparison the pH of a de-ionized water blank is also presented.

Ammonia Volatilization from the Ash

Ammonia volatilization tests were performed by placing thirty-gram samples of hopper ash in a petri dish and a beaker containing 50 mL of 0.1 N sulfuric acid together in individual, sealed, 2.65 L plastic containers. The ammonia concentration of the acid solution was measured to determine the extent of transfer of ammonia from the ash to the liquid. Three trials were needed to produce appropriate results. Initially, it was assumed (incorrectly) that the ammonia on the ash would volatilize continuously. Duplicate samples of ash from each of the three operating conditions (at the three NH_3/NO_x ratios) were placed in six containers, two for each NH_3/NO_x ratio. The first sample for each NH_3/NO_x ratio was analyzed after 14 days had elapsed. It was found that nearly all of the ammonia from the ash had transferred to the acid solution. The second sample was then analyzed two days later, confirming the earlier result. Table 4-4 summarizes these test results. Between 71% and 99% of the original ammonia in the ash volatilized and was captured in the sulfuric acid solution. There appeared to be a dependency of volatilization on the original concentration of ammonia in the ash, however, the NH_3 remaining on the ash was more or less independent of the ammonia initially present on the ash. Ammonia recovery values were 87% or higher.

It should be noted that the samples in the initial closed-container experiment were subjected to wide variations in temperature. Overnight temperatures fell below 30°F in the laboratory when no heat was on over a weekend. Normal indoor temperatures above 70°F prevailed during working hours. At the cooler temperature, the air in the sealed containers became saturated with water vapor and condensation was observed on the interior container walls. When the containers were opened to analyze the acid solution for ammonia, it was also observed that the ash sample had agglomerated.

Table 4-3. Mass concentration, slurry pH, and NH₃ concentration for the isokinetic ash samples and the hopper ash samples used for the SCR ash study.

Sample Code Number	Mass Concentration gr/dscf @ 3% H_2O (dry)	NH_3/NO_x	pH *	NH_3 Conc. µg/g
BI 2	2.93	na	9.44	<6
BI 3	3.01	na	9.03	5
BI 4	3.20	na	7.98	<4
BI 5	3.15	na	7.90	8
BI 6	3.26	na	8.82	<6
BI 7	2.92	na	8.90	<6
BI 8	2.64	na	9.05	6
BAO 2	3.10	0.8	9.94	185
BAO 3	2.89	0.8	8.87	236
BAO 4	3.14	0.8	9.22	175
BAO 5	2.74	0.6	9.21	93
BAO 6	3.13	0.6	9.19	79
BAO 7	2.83	1.0	9.69	888
BAO 8	2.68	1.0	9.85	735
Hopper Ash Sample		0.6	10.17	49
Hopper Ash Sample		0.8	10.10	234
Hopper Ash Sample		1	10.43	349
D.I. H_2O Blank			5.60	

* samples were agitated in sealed bottles for 4 hours before measuring pH (except D.I. H_2O)

The initial attempt to characterize ammonia volatilization from the SCR ash indicated that water absorption and the resulting pH may play an important role in the mechanisms involved. The behavior of the pH of a mixture of SCR ash and deionized water over a period of about 8 minutes is shown in Figure 4-1. Fifty milliliters of deionized water were allowed to come to pH equilibrium with the atmosphere and one gram of ash was then added while the mixture was stirred. The pH of the slurry dropped immediately from 5.3 to 3.7 and then rose rapidly until finally stabilizing at 11.1. The volatilization of the ammonium compounds on the ash to gaseous ammonia would be expected to begin as the pH rose to 9 and above.

The results of a second closed-container volatilization experiment are shown in Table 4-5. Samples of condition 22 ash, prepared as in the first closed-container test, were analyzed after 1, 2, 3, and 7 days. The air temperature in the laboratory was kept near 70 °F. Almost no ammonia was found in the sulfuric acid solution for the first three days (0.1 to 0.2%), however, a significant percentage of the ammonia (87%) had transferred to the acid solution after 7 days. The weight gain of the ash samples, presumably due to absorbed water, was also measured. Almost no ammonia transfer was noted when the weight gain of the 30 g ash sample was 1.6% after three days, however, a significant percentage of the ammonia (87%) had transferred to the acid solution when the weight gain of the ash was only 2.1% after seven days. Ammonia recovery values for this experiment were very good (87.0 to 104.3%).

A third closed-container volatilization experiment was conducted to determine when ammonia transfer from the ash to the acid solution occurred. Seven samples of condition 22 ash were placed in separate containers along with beakers of 1:1 sulfuric acid. Laboratory temperature was kept near 70 °F. A container was opened and the ash and acid solution analyzed for ammonia daily for the succeeding 7 days. Table 4-6 shows the results of these tests. Most of the ammonia was transferred from the solid to the liquid between the fourth and sixth days. As in the earlier experiment, little ammonia transfer was observed until the ash weight had increased by 1.8% or greater. Figure 4-2 shows a plot of the fractions of ammonia on the ash and in the sulfuric acid solution over the seven days of this experiment. This graph shows the rate of transfer of ammonia from one phase (solid) to another (gaseous). The two points from the fourteenth day and the sixteenth day from the earlier work were added to show the continuing nature of the ammonia volatilization.

A dynamic volatilization experiment was also run in which a large volume of ambient air was passed over a sample of ash before being scrubbed of ammonia in a set of impingers filled with 0.1 N sulfuric acid. This experiment with ash in the sample container was run twice. To determine whether ammonia actually volatilized from the ash, a third, blank test was conducted in which the identical sampling procedure was followed, but there was no ash in the sample container. A set of 100 mL impingers, two containing 50 mL of acid and a third serving as a trap, was used for the first test. Due to the small volume of acid and the long duration of the run, the liquid was lost by evaporation and it was necessary to
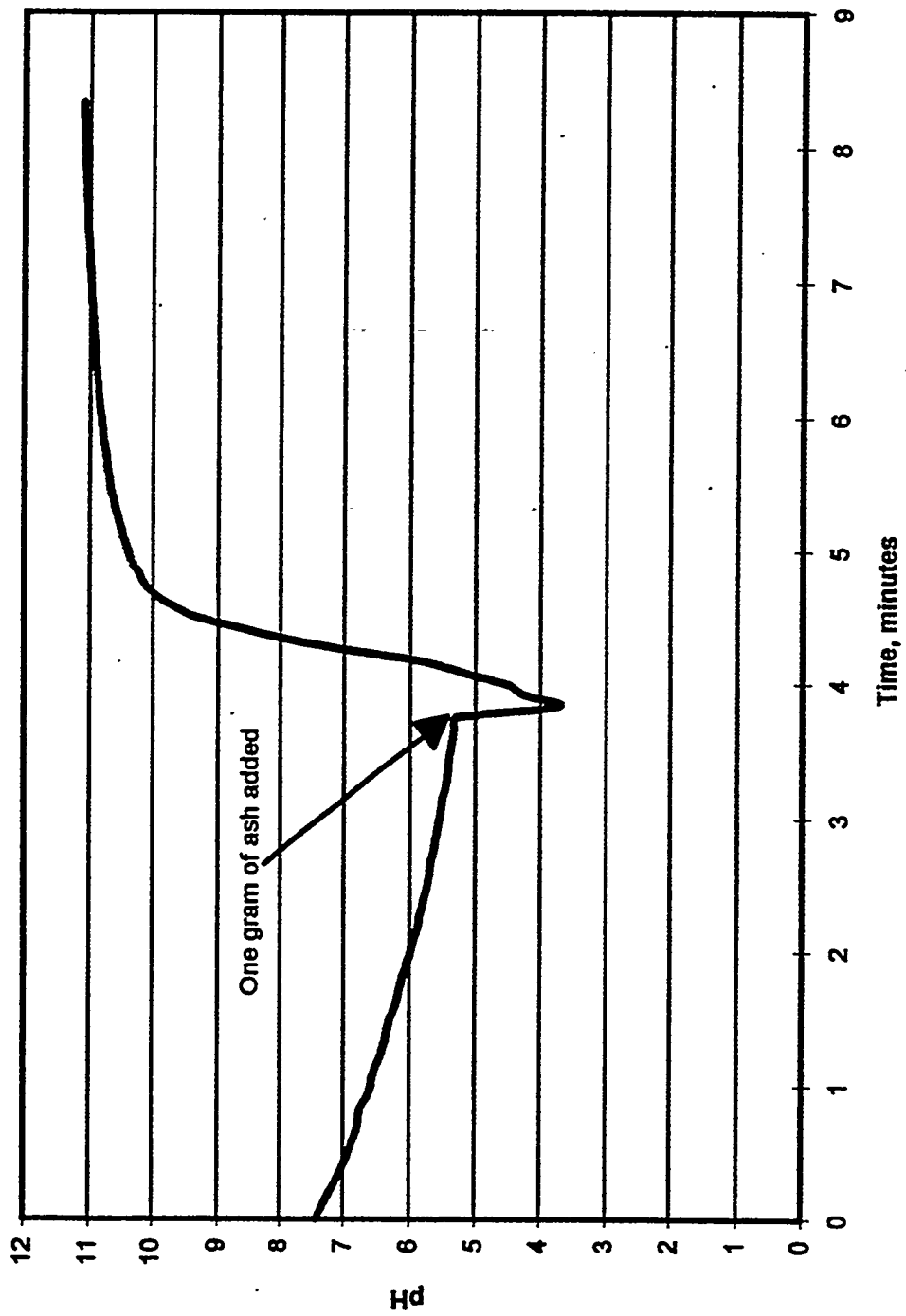
Figure 4-1. pH of a slurry of one gram of ash in fifty milliliters of deionized water.

Table 4-5. Summary of the second closed container volatilization experiment using ash collected at 0.8 NH_3/NO_x ratio.

Test Duration, days	Initial NH_3 On Ash, μg	Final NH_3 On Ash, μg	Ash Mass Gain, %	Final NH_3 In H_2SO_4, μg	NH_3 Volatilized, %	NH_3 Recovery, %
1	7,054	6,965	0.9%	7	0.1%	98.8%
2	7,053	7,344	1.3%	13	0.2%	104.3%
3	7,167	7,144	1.6%	0	0.0%	99.7%
7	7,184	333	2.1%	5,920	82.4%	87.0%

Table 4-6. Summary of the third closed container volatilization experiment using ash collected at 0.8 NH_3/NO_x ratio.

Test Duration, days	Initial NH_3 On Ash, µg	Final NH_3 On Ash, µg	Ash Mass Gain, %	Final NH_3 In H_2SO_4, µg	NH_3 Volatilized, %	NH_3 Recovery, %
1	7,358	7,285	0.8%	< 6	0.1%	99.1%
2	7,355	6,806	1.2%	11	0.2%	92.5%
3	7,325	6,949	1.5%	8	0.1%	94.9%
4	7,325	6,731	1.7%	6	0.1%	92.0%
4	7,425	6,918	1.7%	7	0.1%	93.2%
5	7,322	3,883	1.8%	1,843	26.5%	79.6%
6	7,331	1,258	1.9%	6,058	82.6%	99.8%
7	7,418	490	2.0%	5,768	77.8%	84.4%
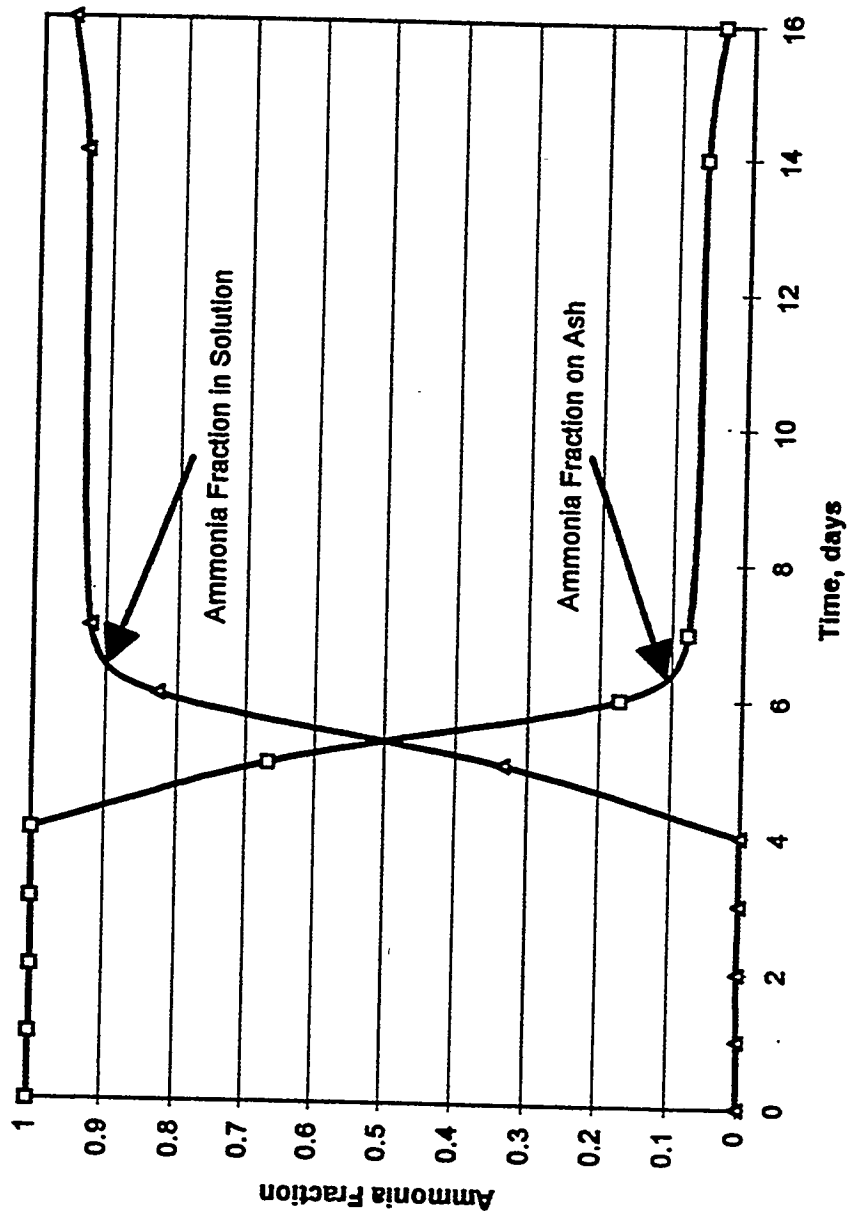
Figure 4-2. Transfer of ammonia from the solid phase to a sulfuric acid solution. Both the ash and the acid solution were in a in a sealed container.

replenish the acid solution periodically during the first run which was terminated after five days (sample 115-NHx-ASH-01). For the second test and the third blank test, three 500-mL Smith-Greenburg impingers, two containing 200 mL of 0.1 N sulfuric acid and the third serving as a trap, were used (sample 115-NHx-ASH-02 and blank 121-NHx-ASH-01). The second test was run for ten days. The blank test was run for eight days.

The analytical results of these three tests are shown in Table 4-7. A small, but measurable, mass of ammonia was collected in the impingers during all three tests, indicating that there was no measurable volatilization of ammonia from the dry ash. The ambient air used for this experiment contained an unknown and unmeasured concentration of water vapor, but it was known to be well below the saturation point (< 100% RH). Thus, in this experiment, the water vapor could not be taken up to reach the nominal 2% level and therefore little NH_3 volatilization occurred. For this test then, volatilization was prevented, not by the dynamics, but by the absence of water vapor.

Ammonia Extraction

The extraction of ammonia in an ash/water slurry was also investigated. A test matrix of nine samples defined by three different quantities of ash and three different pH levels were analyzed for ammonia content. Samples of approximately 1, 2, and 3 grams of the condition 22 hopper ash were placed in fifty milliliters of deionized water, fifty milliliters of an acetate buffer, and in fifty milliliters of a phosphate buffer, allowed to stand for 24 hours, and analyzed for ammonia concentration. The resulting pH levels and ammonia concentrations are shown in Table 4-8. Ammonia concentration of the buffered solutions agreed with predicted concentrations based on ammonia extractions in dilute sulfuric acid. However, the ammonia concentration of the ash samples extracted in deionized water were much lower than predicted.

Two factors may have produced the lower ammonia concentrations observed in the deionized water slurry. First, a portion of the ammonia may not have been extracted from the ash, and secondly, the ammonia may have escaped from the open beakers as gas-phase ammonia. Because gas-phase ammonia is liberated in solutions at higher pH levels like those observed when SCR ash was mixed with water, the water extractions were repeated using sealed bottles. The ammonia concentrations shown in Table 4-9 for the ammonia extractions in sealed bottles with deionized water were higher than when open beakers were used, but still lower than predicted. The ash filtrate was re-extracted in the acetate buffer solution with results as shown in the table. These results indicate that a portion of the ammonia escaped from the solution before the analysis was completed and that another portion was not extracted from the ash by the deionized water.

Table 4-7. Dynamic ammonia volatilization experiment.

Sample	Ash Mass, g	Liquid Sample Volume, ml	Dilution Factor	Conc., µg/ml as N	NH_3 Mass, µg	Air Volume, Standard L	NH_3 Conc. in Air, ppm(v) (dry)
115-NHx-ASH-01	260	113.4	1	0.645	89	23,283	0.005
115-NHx-ASH-02	220	473.8	1	0.201	116	33,927	0.004
121-NHB-ASH-01	0	388.9	1	0.0601	28	12,178	0.003

Table 4-8. Ammonia extractions from Test Condition 22 (0.8 NH₃/NOₓ) hopper ash at three pH values.

Sample	Ash Mass, g	Volume, ml	Solvent	pH	D_F	Conc., µg/ml as N	NH₃ Mass, µg	Apparent NH₃ Concentration on Ash, µg/g
Buffer 1				4.62				
Buffer 2				6.18				
1	1.0155	50	Buffer 1	4.68	5	0.814	247	243
1D	1.0526	50	Buffer 1	4.67	5	0.852	259	246
2	3.0071	50	Buffer 1	4.67	10	1.19	723	240
2D	3.1297	50	Buffer 1	4.67	10	1.26	765	244
3	5.1103	50	Buffer 1	4.67	10	2.05	1245	244
3D	5.2750	50	Buffer 1	4.67	10	2.19	1330	252
4	1.0423	50	Buffer 2	6.16	10	0.485	294	283
4D	1.1374	50	Buffer 2	6.16	10	0.465	282	248
5	3.0744	50	Buffer 2	6.17	10	1.15	698	227
5D	3.0890	50	Buffer 2	6.17	10	1.19	723	234
6	5.0583	50	Buffer 2	6.18	10	1.66	1008	199
6D	5.0772	50	Buffer 2	6.19	10	1.84	1117	220
7	1.0153	50	D.I. H₂O	9.10.	1	1.69	103	101
7D	1.0442	50	D.I. H₂O	9.19	1	1.68	102	98
8	3.0233	50	D.I. H₂O	9.51	1	3.3	200	66
8D	3.0824	50	D.I. H₂O	9.54	1	3.3	200	65
9	5.0525	50	D.I. H₂O	9.75	1	4.05	246	49
9D	5.0933	50	D.I. H₂O	9.67	1	6.39	388	76
3D spike	(+2.0 µg/ml)					4.14		
6D spike	(+2.0 µg/ml)					3.54		
1.00 STD						1.06		
D.I. H₂O blank						0.013/BDL		

Table 4-9. Closed-container ammonia extractions from Test Condition 22 hopper ash in D.I. water.

Sample	Ash Mass, g	Volume, ml	Solvent	pH	D_F	Conc., µg/ml as N	NH₃ Mass, µg	Apparent NH₃ Concentration on Ash, µg/g
10	1.013	111.99	D.I. H₂O	11.06	2	0.693	188	186
11	3.020	93.76	D.I. H₂O	11.31	2	1.74	396	131
12	5.000	96.38	D.I. H₂O	11.45	2	3.44	805	161
Re-extractions in acetate buffer, pH 4.8								
10	1.013	37.0	4.8	na	5	0.249	58	55
11	3.020	42.9	4.8	na	5	0.254	66	22
12	5.000	41.1	4.8	na	5	0.306	76	15
Extraction of ash in 0.1 N sulfuric acid								
Ash	1.004	42.5			25	0.232	299	298

Particle Size Dependency of Ammonia Concentration

To investigate the dependency of ammonia concentration on particle size, two special tests were conducted at the outlet of the Reactor B air heater with a five-stage cyclone particle size instrument. This device which is operated in situ samples the flue gas isokinetically and then passes it though five sequential cyclones. Each cyclone is designed to remove a smaller size particle. For this test the cyclone "cut" sizes were approximately 7.8, 4.5, 2.5, 1.8, and 0.8 micrometers diameter (Stokes aerodynamic).

Tables 4-10 and 4-11 show data summaries for the two cyclone tests at the outlet of the Reactor B air heater. The analysis of the distribution of ammonia was confined to the ash actually captured in the five cyclone stages since a quantitative analysis of pre-collector and back-up filter material was not conducted. For the purpose of this study, the median particle diameter for each stage, the mass of ash collected in each stage, and the mass-basis ammonia concentration of the particulate in each stage were of primary interest. The data in the tables confirmed that the ammonia concentration (mass basis) of the fly ash is strongly dependent on particle size. The ammonia concentration in the ash increased by nearly two orders of magnitude from the 7.8 micrometer diameter particles to the 0.8 micrometer diameter particles. This relationship is illustrated in Figure 4-3. A power curve is superimposed on the data to illustrate the apparent trend of the relation but is not meant to suggest any definitive relationship.

The ammonia concentration in the ash was much higher in the smaller particle sizes, but most of the total ammonia was found to reside with the larger particles simply because these comprise the vast majority of the ash mass. The ammonia distribution is presented as a fraction of the total ammonia resident in each particle size group in Figure 4-4. The linear trend line superimposed on the data is included only to aid the visual presentation of the data and is not meant to show that any relationship has been conclusively demonstrated. The average total mass of ash collected for the two cyclone runs was 6.026 grams and the average mass of ammonia extracted form all the ash was 528.7 micrograms. These values give a ammonia concentration (mass basis) of 88 ppm(w) or micrograms of ammonia/gram of ash.

Figure 4-5 shows the cumulative ammonia distribution plot of the sum of all ammonia in the particle size at which the point is plotted plus all of the ammonia in the smaller particle size stages. These data imply that very little slip ammonia will exit the stack when high efficiency particulate emission controls are in place since all detectable ammonia is in the solid phase at the air heater exit and most of the ammonia is associated with the larger particle sizes which are most readily collected.

Table 4-10. Summary of the first five-stage cyclone test.

TEST DESIGNATOR: Cyclone 1
TEST TYPE: INLET
RUN NUMBER: 128-BAO-CYC-01

Actual Flow Rate: 0.889 ft³/min
Standard Flow Rate: 0.532 ft³/min
% Isokinetic: 96.57 %
Viscosity: 229.3E-06 g/g-s
Calculated Cyclone ΔP: 1.92 in. Hg

Stage	Cunningham Correction Factor	D_{50}, µm (Classical Aerodynamic)	D_{50}, µm (Stokes Aerodynamic)	Cumulative Frequency, %	Reynolds Number	$\sqrt{\psi}_{50}$	Mass, g	NH_3 Conc., µg/g
1	1.033	7.639	7.765	15.7444	1461	0.195	5.1647	24.6
2	1.059	4.277	4.402	5.0997	1858	0.158	0.6525	98.8
3	1.109	2.333	2.456	1.7113	2475	0.138	0.2077	198.2
4	1.158	1.616	1.739	0.3051	3639	0.172	0.0882	421.4
5	1.401	0.677	0.802	0.2007	6085	0.172	0.0064	734.0
Back-Up Filter							0.0123	

Stage Cut Diameters Based on Particle Density = 2.5 g/cm³

Total Mass Concentration:
  6780 mg/m³ (dry), STD
  4060 mg/m³ actual
  2.96 gr/scf, dry
  1.77 gr/acf

Table 4-11. Summary of the second five-stage cyclone test.

TEST DESIGNATOR: Cyclone 2
TEST TYPE: INLET
RUN NUMBER: 128-BAO-CYC-02

Actual Flow Rate:	0.866	ft³/min
Standard Flow Rate:	0.516	ft³/min
% Isokinetic:	93.78	%
Viscosity:	229.3E-06	g/g-s
Calculated Cyclone ΔP:	1.82	In. Hg

Stage	Cunningham Correction Factor	D_{50}, μm (Classical Aerodynamic)	D_{50}, μm (Stokes Aerodynamic)	Cumulative Frequency, %	Reynolds Number	$\sqrt{\psi_{50}}$	Mass, g	NH_3 Conc., μg/g
1	1.033	7.781	7.907	15.5336	1423	0.196	5.0951	22.8
2	1.058	4.393	4.518	5.0065	1807	0.160	0.6350	104.3
3	1.106	2.399	2.522	1.7606	2409	0.138	0.1958	189.7
4	1.154	1.660	1.783	0.3597	3543	0.174	0.0845	403.3
5	1.387	0.698	0.822	0.2039	5924	0.174	0.0094	734.0
Back-Up Filter							0.0652	

Stage Cut Diameters Based on Particle Density   =   2.5 g/cm³

Total Mass Concentration:

6880	mg/m³ (dry), STD
4100	mg/m³ actual
3.01	gr/scf, dry
1.79	gr/acf
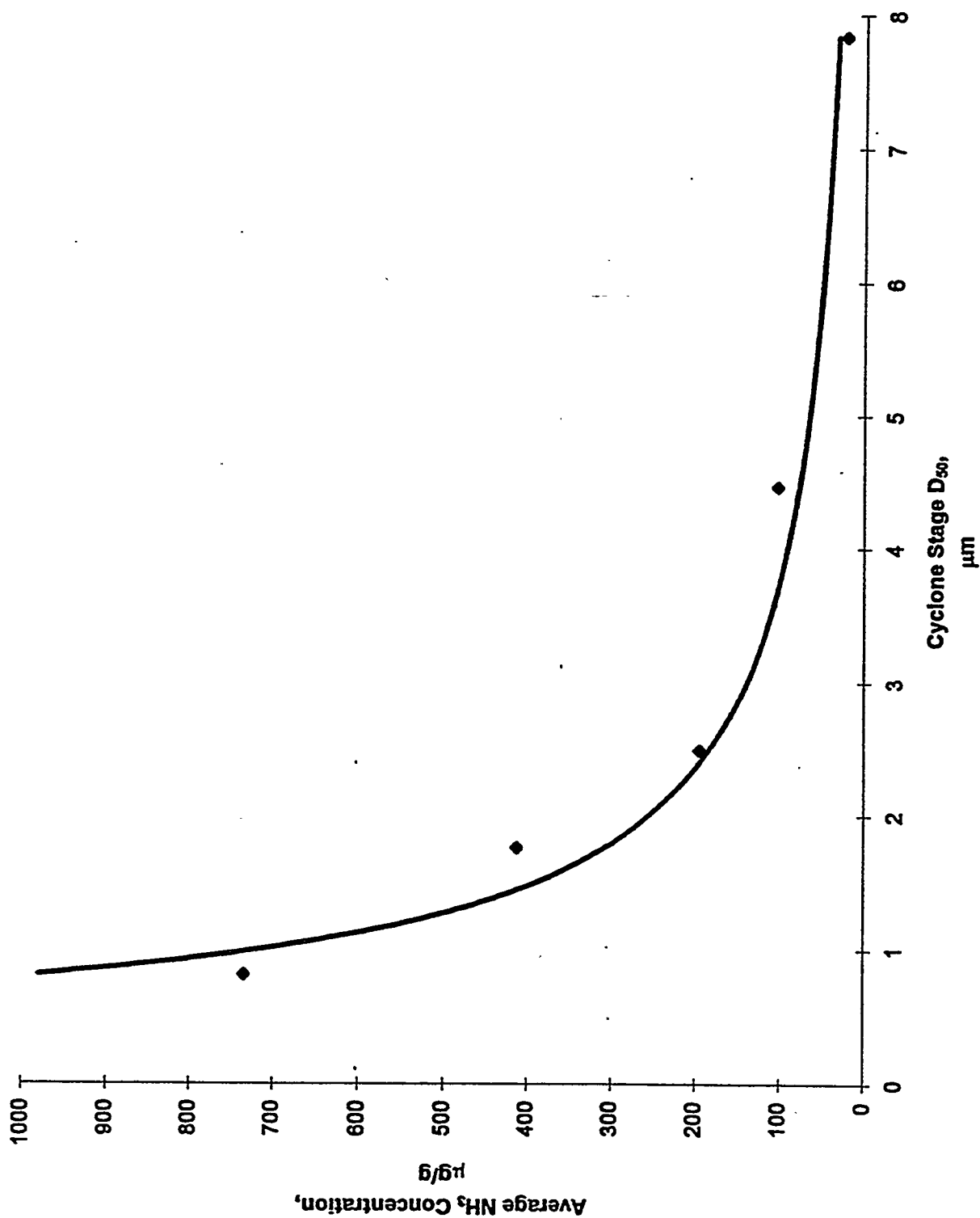
Figure 4-3. Mass concentration of ammonia as a function of particle size.
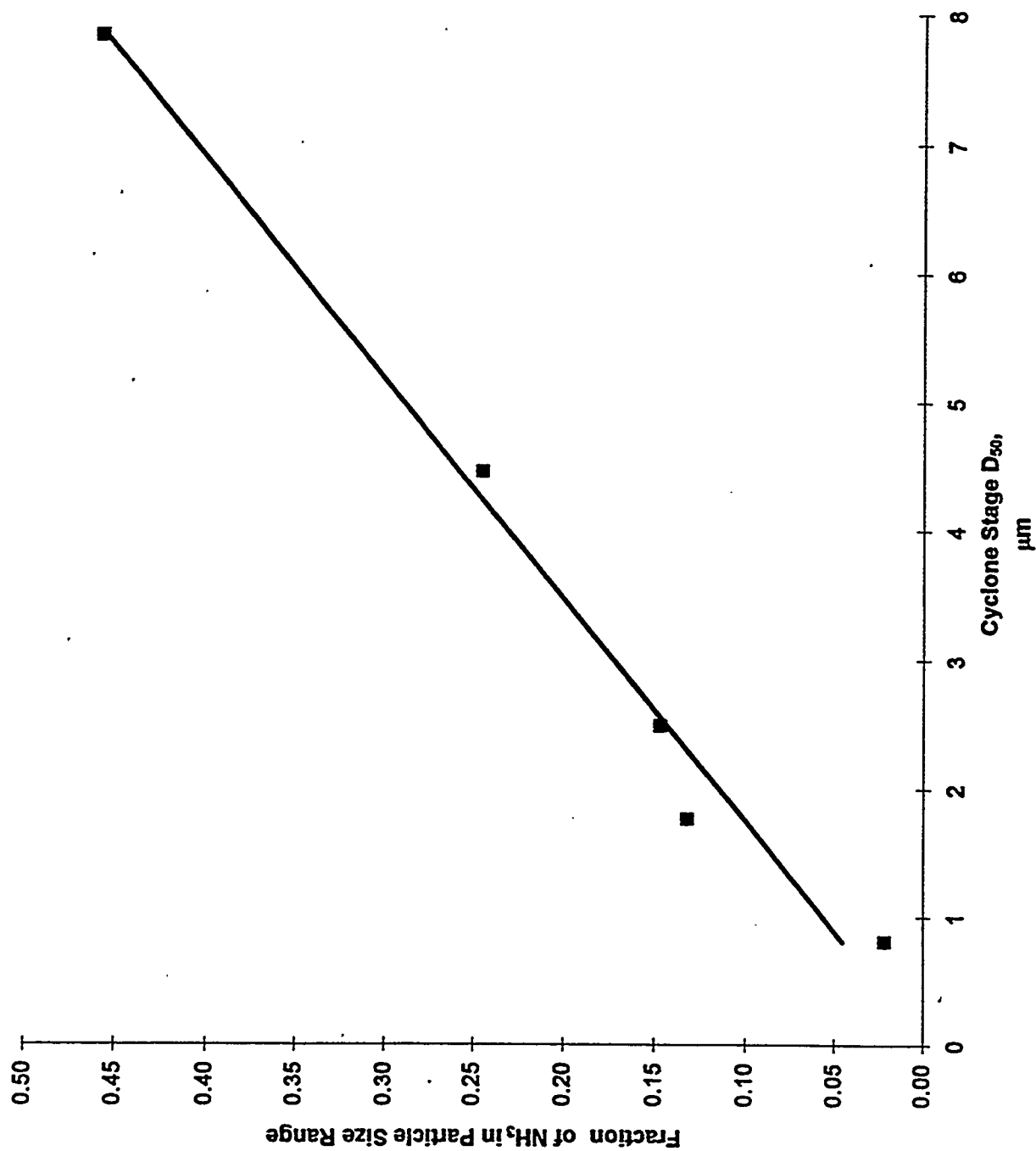
Figure 4-4. Distribution of ammonia among the five particle size ranges.
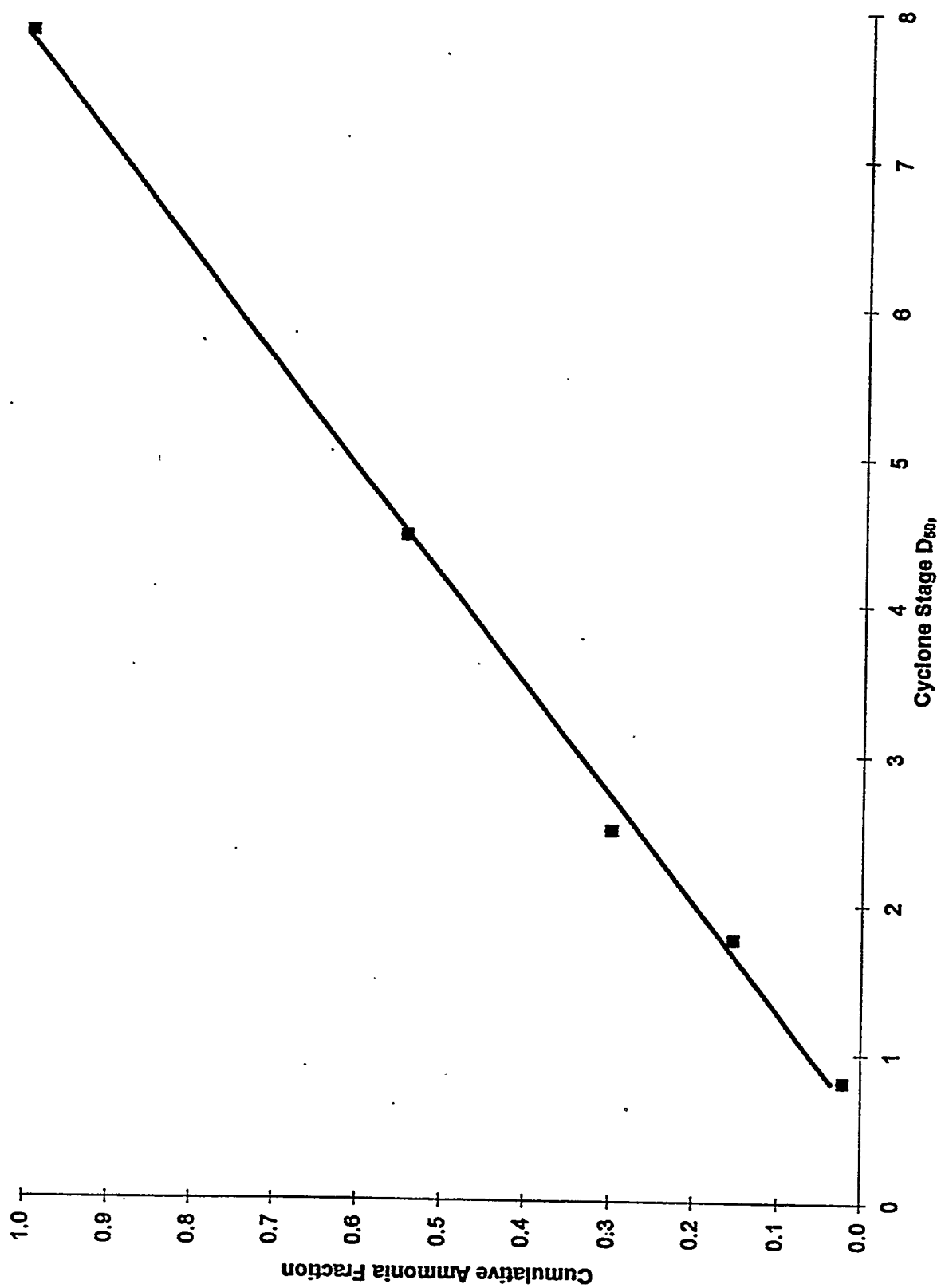
Figure 4-5. Cumulative solid-phase ammonia fraction as a function of particle size.

Conclusions

Almost no ammonia volatilized from the SCR ash until a significant amount of water vapor was absorbed by the ash. A plausible mechanism for the apparent volatilization that occurred is that enough water was gained by the ash to form a moist layer with a pH high enough to evolve gas-phase ammonia from the ammonium compounds on the ash. Nearly all of the ammonia on the ash evolved to the gas phase in the closed-container experiments. Ammonia concentrations in enclosed spaces will depend on the ammonia concentration of the ash, the volume of air above the ash available for dilution, and the presence of a humid atmosphere.

The extraction of ammonia from fly ash seems to depend upon pH. Evidently, all or nearly all, of the ammonia present was extracted in the buffered solutions at pH 4.7 and pH 6.2, but not all was recovered in alkaline unbuffered extracts. In the pH 6.2 buffer, however, the completeness of extraction seemed to fall off somewhat as the ratio of ash to buffer increased. At 3 g of ash per 50 mL of pH 6.2 buffer, the amount of ammonia extracted was about 200 μg/g, whereas at 1 g per 50 mL, the amount was near 250 μg/g.

The hopper ash that was collected at the same NH_3/NO_x ratio as the suspended fly ash (ratio, 0.8) was extracted with sulfuric acid (at pH 1.7) and found to contain ammonia at a concentration of about 250 μg/g, or about the same concentration as in the fly ash extracted with pH 4.7 buffer. The ashes differ in particle size and perhaps in ammonia content; if the hopper ash contains less ammonia, as may reasonably be suspected, because of its coarser particle size, then it is conceivable that the equivalent amounts of ammonia extracted from both ashes mean the pH 4.7 buffer is somewhat less effective for extraction than the dilute sulfuric acid. The available data do not permit this uncertainty to be fully resolved.

The effect of pH described above indicates that extraction is more complete when ammonia appears in the extract as the ammonium ion NH_4^+ rather than the free base molecule NH_3. At pH 1.7, the ratio NH_4^+/NH_3 is approximately 10,000,000. At pH 4.7, the value decreases to about 10,000, and at pH 6.2 it is roughly 100. When the pH reaches 11, however, in the absence of a buffer, NH_3 is predominant, and the NH_4^+ ratio is only about 0.01. The correlation of extraction with NH_4^+ ion, however, cannot be explained theoretically. The solubility of NH_3 is not limited in a practical way by pH. The volatility of NH_3, on the other hand, is appreciable, and this property of NH_3 may be substantially responsible for the data indicating that NH_3 is not completely extracted.

It is surprising to see that the addition of ammonia apparently makes the ash more alkaline (shown by comparing outlet ash with inlet ash). It is probably not the ammonia per se, however, that is responsible. Even without ammonia, the ash makes such a high pH in a slurry that ammonia can contribute nothing as a base. Near pH 11, dissolved ammonia is too weak as a base to capture protons and thus exhibit the property of a base. Ammonia can capture protons when the pH is well below 9 and the proton concentration is relatively

high. It cannot capture protons when the pH is much above 9 and the proton concentration is very low. In such an alkaline medium, the NH_3 molecule is not reactive; its behavior is not very different from that of the water molecule.

The paradox of increased slurry pH is likely due to an indirect effect from ammonia. Ammonia will neutralize the sulfuric acid that would otherwise occur on the ash surface. If sulfuric acid is neutralized by ammonia, it is not available to neutralize other extractable base in the ash and thus the pH of the slurry rises to a higher value.

Section 5

INVESTIGATION OF METALS EXTRACTABILITY FROM FLY ASH

The fourteen, isokinetically collected ash samples were received in our Birmingham laboratories and were composited to produce six samples, one representing each of three chosen NH_3/NO_x molar ratios in the reactor at each of two sampling locations, one at the inlet to SCR Reactor B and one at the outlet from the Reactor B air heater. Note that, because the upstream (inlet) samples were never exposed to ammonia in the reactor, all three of the inlet samples should have been essentially identical in composition. The target NH_3/NO_x molar ratios selected for this study were (nominally) 0.6, 0.8, and 1.0.

Through informal laboratory experimentation with another similar fly ash sample, we learned that the ash produces a solution pH of about 11 when combined with water in the ratio indicated above. We also learned that the large reservoir of soluble base in the ash makes pH control at some other lower value impractical, as painstaking incremental additions of acid would have to be continued indefinitely to achieve a predefined lower pH level. Of course, a buffer could be used to control pH, but the buffer itself might easily alter the extractabilities of certain metals, leading to erroneous conclusions about what might happen in an ash pond. In addition, the wastewater permits that we are aware of limit discharge pHs to the alkaline range. It was primarily for these reasons that, except for one specially prepared ash sample (0.8 NH_3/NO_x molar ratio), we chose to forego any pH adjustment in these sample extracts.

The purpose in lowering the pH of one of the samples was to give insight to the probable direction of extractability changes when factors in the natural environment overcome the high alkalinity of this ash. As noted below, the time interval at this lower pH value was transient indicating the reserve of slowly dissolving alkali in this ash.

Each sample was subjected to an extraction process that was intended to at least roughly simulate the leaching that occurs when ash is sluiced or allowed to settle in an ash pond. That is, duplicate 5-g portions of each sample were accurately weighed into precleaned 125-mL polyethylene (Nalgene) bottles and combined with 100-mL aliquots of deionized water. An extra sample of reactor outlet ash corresponding to a NH_3/NO_x ratio of 0.8 was similarly prepared for use in an experiment involving pH adjustment of the extract. Two extraction blanks were also prepared by adding 100 mL of deionized water to each of two empty bottles. Note that the above amounts of water and ash represent a water-to-ash ratio of 20:1, which is a consistent, but conservative, ratio identified by the customer for sluicing operations. This ratio was also fixed at the lowest reasonable value (consistent with the 20-to-25-g sample sizes that were available) to allow us to detect as many of the metals of interest as possible in the extracts.

All sample and blank solutions were placed in a tumbling apparatus and tumbled for approximately 96 hr. About 24 hr before the end of this period, however, the extra reactor outlet sample mentioned above was briefly removed from the tumbling apparatus and combined with 0.85 mL of 1.0 N sulfuric acid. The sample pH immediately dropped to 7, then began to slowly increase. The Teflon-coated magnetic stir bar added to this sample to facilitate the pH adjustment could not be removed from the solution without losing part of the ash sample. Thus, the stir bar was left in place, and another stir bar was added to one of the blanks to enable it to serve as a blank for this sample. Then all samples were returned to the tumbling apparatus, and tumbling was continued for the duration of the 96-hr agitation interval.

After the above extraction process, each sample was removed from the tumbling apparatus and allowed to stand for about three days to permit the undissolved ash to settle out of the solution. We then checked the solution pH with a pH meter and filtered the aqueous phase through an Acrodisc syringe filter into a precleaned 125-mL polyethylene (Nalgene) bottle. From each filtered sample, a 25-mL aliquot was transferred to a precleaned 60-mL amber glass bottle for mercury determination. The remainder was reserved for microwave digestion and analysis for other metals. The pertinent sample-preparation data are synopsized in Table 5-1.

Preparation of Fly-Ash Extracts for Analysis

The sample extracts were ultimately analyzed for all metals that were thought to be pertinent to the problem at hand, including the 20 metals that were previously determined in Plant Crist fly ash: arsenic (As), barium (Ba), beryllium (Be), cadmium (Cd), cerium (Ce), cobalt (Co), chromium (Cr), copper (Cu), mercury (Hg), manganese (Mn), molybdenum (Mo), nickel (Ni), lead (Pb), rubidium (Rb), antimony (Sb), selenium (Se), tin (Sn), strontium (Sr), vanadium (V), and zinc (Zn). In addition, we also analyzed the extracts for silver (Ag), boron (B), phosphorus (P), thallium (Tl), aluminum (Al), calcium (Ca), iron (Fe), and magnesium (Mg).

The sample extracts were prepared for Hg analysis essentially by the preparation method given in EPA Method 7470. Using this method as a guideline, we diluted a 10-mL aliquot of each extract to 100 mL. The diluted solution was transferred to a BOD (biological oxygen demand) bottle and combined with 5 mL of concentrated sulfuric acid, 2.5 mL of concentrated nitric acid, 15 mL of 5% (w/v) aqueous potassium permanganate, and 8 mL of 5% (w/v) potassium persulfate. The bottle was then placed on a hot plate at 95 °C for 2 hr. Afterward, the sample solution was cooled and combined with 6 mL of an aqueous reagent solution consisting of 12% (w/v) hydroxylamine hydrochloride and 12% (w/v) sodium chloride.

For the purpose of matrix matching, the calibration standards were treated in the same manner as set forth above. A digestion blank and an independent QC sample (i.e., water

Table 5-1. Sample Preparation Data for Plant Crist SCR Fly Ash Samples

Reactor B Inlet Samples

Sample No.	NH$_3$/NO$_x$	Ash Weight, g	Extract Volume, mL	Final pH
1A-1	0.8	5.0092	100	11.1
1A-2	0.8	4.9848	100	11.1
2A-1	0.6	5.0093	100	11.3
2A-2	0.6	5.0161	100	11.2
3A-1	1.0	5.0080	100	11.3
3A-2	1.0	5.0022	100	11.3

Reactor B Air Heater Outlet Samples

Sample No.	NH$_3$/NO$_x$	Ash Weight, g	Extract Volume, mL	Final pH
1B-1	0.8	5.0139	100	11.3
1B-2	0.8	5.0028	100	11.3
1B-3*	0.8	5.0026	100.85	9.1
2B-1	0.6	5.0012	100	11.4
2B-2	0.6	5.0062	100	11.3
3B-1	1.0	5.0071	100	11.7
3B-2	1.0	5.0171	100	11.7

Extraction Blanks

Sample No.	NH$_3$/NO$_x$	Ash Weight, g	Extract Volume, mL	Final pH
WB-1	------	------	100	5.5
WB-2	------	------	100	5.5

*This sample received 0.85 mL of 1.0 N sulfuric acid in an attempt to adjust its pH downward (see text).

containing spiked mercury at a level that was unknown to the analyst) were also prepared in this way. Furthermore, two additional aliquots of one of the real samples were spiked with mercury and carried through the above preparation procedure exactly like the other samples. Real samples spiked and treated in this manner are referred to subsequently in this report as "matrix spikes".

For determinations of all other metals except Rb, the sample extracts were digested according to EPA Method 3015 - Microwave Assisted Acid Digestions for Aqueous Samples and Extracts. That is, a 40-mL aliquot of each sample was placed into a Teflon microwave digestion vessel. After the addition of 5 mL of concentrated nitric acid to each vessel, the vessels were capped and placed in the microwave oven. The samples were heated for a total of 20 min in a two-stage heating program. After a brief cooling period, 0.5 mL of an internal standard solution (scandium, Sc) was added to each vessel, and the contents of each vessel were rinsed into a 50-mL polyethylene (Nalgene) volumetric flask. After mixing the solutions, we transferred them to 125-mL Nalgene bottles.

Two separate digestions of the type described above were required to prepare all of the samples. Thus, each digestion run included a digestion blank. In addition, a certified QC sample was included in one of the runs. Moreover, two matrix spikes were prepared as described previously, i.e., by spiking a real sample with the elements of interest. The spike levels were chosen to be roughly the same as the natural (i.e., unspiked) metal concentrations in the sample, as determined in a preliminary analysis of the sample in question. If the natural metal concentration was at or below our detection limit, then the sample was spiked with that metal at a concentration that was very close to the detection limit to enable us to identify even minor spectral and matrix interferences. All metals of interest were included in the matrix spikes except Ca and B, which were present at such high levels that a comparable spike would have required an unacceptable expansion of the solution volume. For all analyses, calibration standards were made up in 5% (v/v) nitric acid from certified 1000-ppm stock solutions.

For determinations of Rb, the sample extracts were not digested; they were merely combined with concentrated nitric acid to form a 5% (v/v) solution and further combined with potassium chloride to form a 0.2% (w/v) solution.

Analysis of Sample Extracts

All metals except Hg, Pb, Tl, and Rb were initially determined by inductively-coupled plasma atomic emission spectrometry (ICPAES). The ICPAES conditions and methods were essentially those set forth in EPA Method 6010. The instrument was a Perkin-Elmer Plasma 400 Spectrometer with QC Expert software and an AS-90 Autosampler. A six-point calibration curve (including a calibration blank) was used for each element except Ag, which tends to fall out of solution at high concentrations when other metals are present at high concentrations. For Ag, therefore, we used a five-point calibration curve

covering a relatively short concentration range. A calibration check standard was analyzed at the beginning and end of each instrumental run. To check for spectral interferences, As, Sb, and Se were analyzed at each of two wavelengths. All quantitation was performed by computing the ratio of the response from the analyte to the response from the internal standard (Sc).

Graphite-furnace atomic absorption spectrometry (GFAAS) was the only analysis method used for determining Pb and Tl. In addition, after we reviewed the ICPAES data, we decided to use this technique also to redetermine Ag, As, Cd, Cu, and Sb, mainly in an attempt to achieve lower detection limits. A calibration check standard and two calibration blanks were run essentially as described above for ICPAES. A 20-ppm Ni solution was used as a matrix modifier for the Tl, As, and Sb analyses. Our GFAAS methods were adapted from the 7000 series of methods promulgated in EPA SW-846. The instrument used for this work was a Perkin-Elmer Model 3100 Atomic Absorption Spectrometer equipped with an HGA-600 Graphite Furnace assembly, an AS-60 Autosampler, and Perkin-Elmer 5100 Software. The instrument uses deuterium background correction, electrodeless discharge lamps (for As and Sb), hollow cathode lamps (for all other elements), and a pyrocoated graphite tube fitted with a L'Vov platform.

After a review of the GFAAS data for Cu, we decided that the ICPAES method had provided a slightly better detection limit; hence, the Cu data reported here are those from the ICPAES analyses. But for those samples that yielded an unambiguously positive ICPAES response to Cu, we were able to confirm the found concentrations from the GFAAS data.

Our Hg analyzer, a PSA Analytical Merlin-Plus System with a hydride/vapor generator and an autosampler, has the unique feature of analyzing samples simultaneously by two different methods: atomic absorption and atomic fluorescence. The atomic fluorescence unit is the more sensitive one; it is able to measure Hg concentrations down to about 20 parts per trillion in calibration standards. With regard to calibration check standards and other QC samples, this instrument was evaluated during use in much the same way as the others.

Determinations of Rb were carried out by the atomic emission technique on a Perkin-Elmer Model 2380 instrument. A slot-type air-acetylene flame was used as the emission source. Again, QC checks were much the same as for the other instruments.

In this test program, we used an aggressive but risky definition of detection limit. Specifically, we chose to report numerical concentration values for all analyte responses lying more than one standard deviation above the average response to two digestion blanks, two calibration blanks, and two extraction blanks. (There were generally no significant differences in response from these three different types of blanks.) In a table of normal curve areas, one finds that about 14% of blank responses can be expected to lie at

or beyond one standard deviation above the average blank value, assuming that the blank data at least roughly approximate a normal distribution. Thus, the risks associated with this approach, i.e., the risk of mistaking a blank response for an analyte signal (and vice versa), are quite significant.

However, this approach also maximizes the useful quantitative information obtainable from those analyses where the analyte is present at or near the detection limit. Note that, if the odds are about 14% that one blank measurement will fall at or above the one-standard-deviation cutoff, then the odds are substantially less than 14% that two successive blank measurements will fall at or above this level. Thus, one can examine the results of two or more replicate analyses and often make reasonably sound, low-risk judgments about the validity of response data that might otherwise have been discarded because they were too close to the detection limit. Nevertheless, all response data lying between one and three standard deviations above the average blank response are, in this report, enclosed within parentheses to highlight their tenuous nature.

Test Results

Table 5-2 contains the chemical analysis data for the Plant Crist fly-ash sample extracts. Note that the table reflects duplicate analyses (Replicate No. 1 and Replicate No. 2) of each sample extract. In cases where the response to a particular metal could not be distinguished from the blank response, the found concentration was reported as "less than" (<) the detection limit.

From the data of Table 5-2, we concluded that Ag, Be, Ce, Cu, Hg, Mn, Ni, Pb, Sn, Tl, and Fe were present at such low levels in most of the sample extracts that they could not be determined with any confidence. Although some samples produced detectable amounts of some of these elements, most samples yielded either no detectable response or a very weak response, so that no conclusions could be drawn with regard to the effects of ammonia on the extractabilities of these metals. Hence, all statements in this report about the effects of ammonia on the extractabilities of the metals should be interpreted to pertain only to the detectable metals in our study.

We compared the data of Table 5-2 with the results of previously obtained analyses of two fly-ash samples from Plant Crist with a view toward roughly estimating the fraction of each metal extracted in our experiments. We found that the fraction of the metal content extracted from the ash in the present study ranged from about 0.1 down to <0.001 for the majority of metals for which fly-ash analysis data were available (i.e., all metals of interest here except Ag, B, Ce, P, Al, Ca, Fe, and Mg). Very high extractabilities, roughly on the order of 0.5 (50%) to 1.5 (150%), were estimated for Mo and Se. This result for Se may be explained on the basis of the very high volatility of selenium dioxide at the temperature of the reactor inlet, but much reduced volatility at the air heater outlet.. That is, most of the Se in the fly ash at the air heater outlet might reasonably be expected to be present as a

Table 5-2. Metals Concentrations in Extracts of SCR Fly Ash Samples

Concentration, mg/mL (ppb)

Element	NH3/NOx	Reactor Inlet			Reactor Air Heater Outlet			Low pH Sample[b]
		Replicate #1[a]	Replicate #2[a]	Average	Replicate #1[a]	Replicate #2[a]	Average	
Ag	0.6	<2.5	<2.5	<2.5	<2.5	<2.5	<2.5	
	0.8	<2.5	<2.5	<2.5	<2.5	<2.5	<2.5	<2.5
	1	<2.5	<2.5	<2.5	<2.5	<2.5	<2.5	
As	0.6	104	101	103	68.6	64.9	66.8	
	0.8	86.9	84.4	85.6	67.3	60.5	63.9	98.8
	1	93.6	90.8	92.2	53	49	51	
B[c]	0.6	15	15.1	15.1	10.8	11.9	11.4	
	0.8	17	18.9	18	12.5	14	13.3	17.9
	1	16.3	18.3	17.3	11.3	12.5	11.9	
Ba	0.6	192	191	192	211	238	225	
	0.8	198	215	207	226	264	245	142
	1	200	241	221	271	294	283	
Be	0.6	<5.09	(5.25)	<5.25	<5.09	6.05	<5.57	
	0.8	<5.09	(5.25)	<5.19	<5.09	<5.09	<5.09	5.66
	1	<5.09	<5.09	<5.09	<5.09	<5.09	<5.09	
Cd	0.6	1.32	0.488	0.904	<0.250	<0.250	<0.250	
	0.8	0.712	3.68	2.2	2.24	<0.250	<1.25	1.4
	1	0.55	0.475	0.513	<0.250	<0.250	<0.250	
Ce	0.6	<46.2	<46.2	<46.2	<46.2	<46.2	<46.2	
	0.8	<46.2	<46.2	<46.2	<46.2	(50.2)	<48.2	<46.2
	1	<46.2	<46.2	<46.2	<46.2	<46.2	<46.2	
Co	0.6	(6.44)	<5.76	<6.10	(9.91)	(12.2)	11.1	
	0.8	(10.2)	(11.4)	10.8	<5.76	13.9	<9.83	<5.76
	1	<5.76	(8.23)	<7.00	(11.4)	(12.8)	12.1	
Cr	0.6	74.9	66.5	70.7	59.3	60.6	60	
	0.8	84.8	82.8	83.8	81.5	77.6	79.6	69.1
	1	109	91.9	100	84.1	73	78.6	
Cu	0.6	<3.21	<3.21	<3.21	(3.40)	(7.08)	5.24	
	0.8	<3.21	<3.21	<3.21	12.4	<3.21	<7.81	4.91
	1	(4.66)	(3.53)	4.1	<3.21	<3.21	<3.21	

Table 5-2. Metals Concentrations in Extracts of SCR Fly Ash Samples, continued

Concentration, mg/mL (ppb)

Element	NH3/NOx	Reactor Inlet			Reactor Air Heater Outlet			
		Replicate #1[a]	Replicate #2[a]	Average	Replicate #1[a]	Replicate #2[a]	Average	Low pH Sample[b]
Hg	0.6	<0.200	<0.200	<0.200	<0.200	<0.200	<0.200	
	0.8	<0.200	<0.200	<0.200	<0.200	<0.200	<0.200	<0.200
	1	<0.200	<0.200	<0.200	<0.200	<0.200	<0.200	
Mn	0.6	<0.240	<0.240	<0.240	<0.240	<0.240	<0.240	
	0.8	<0.240	<0.240	<0.240	(0.28)	<0.240	<0.26	2.63
	1	3.43	<0.240	<1.84	<0.240	<0.240	<0.240	
Mo	0.6	722	743	733	737	798	867	
	0.8	810	866	838	745	844	795	880
	1	771	877	824	758	835	797	
Ni	0.6	<6.60	<6.60	<6.60	(7.79)	(8.24)	8.02	
	0.8	<6.60	<6.60	<6.60	68	<6.60	<37.3	9.75
	1	(13.0)	<6.60	<9.80	(9.80)	<6.60	<8.20	
P	0.6	(98.8)	(79.6)	89.2	<76.1	<76.1	<76.1	
	0.8	(105)	(95.2)	100	<76.1	<76.1	<76.1	92.5
	1	<76.1	(77.1)	<76.6	<76.1	<76.1	<76.1	
Pb	0.6	<6.25	<6.25	<6.25	<6.25	<6.25	<6.25	
	0.8	<6.25	<6.25	<6.25	<6.25	<6.25	<6.25	<6.25
	1	<6.25	<6.25	<6.25	<6.25	<6.25	<6.25	
Rb	0.6	140	140	140	120	120	120	
	0.8	140	140	140	100	100	100	100
	1	120	120	120	80	80	80	
Sb	0.6	116	103	110	65.1	57.5	61.3	
	0.8	116	124	120	47.6	54.5	51.1	152
	1	97.4	89.6	93.5	33	30.6	31.8	
Se	0.6	<48.8	<48.8	<48.8	244	228	236	
	0.8	<48.8	<48.8	<48.8	469	518	494	516
	1	<48.8	<48.8	<48.8	350	390	370	
Sn	0.6	<22.0	(24.0)	<23.0	<22.0	<22.0	<22.0	
	0.8	<22.0	<22.0	<22.0	<22.0	<22.0	<22.0	(22.5)
	1	<22.0	<22.0	<22.0	<22.0	<22.0	<22.0	

Table 5-2. Metals Concentrations in Extracts of SCR Fly Ash Samples, continued

Concentration, mg/mL (ppb)

Element	NH3/NOx	Reactor Inlet Replicate #1[a]	Replicate #2[a]	Average	Reactor Air Heater Outlet Replicate #1[a]	Replicate #2[a]	Average	Low pH Sample[b]
Sr[c]	0.6	1.02	1.14	1.08	0.974	1.16	1.07	
	0.8	1.03	1.16	1.1	0.965	1.07	1.02	1.07
	1	0.954	1.06	1.01	0.878	0.945	0.912	
Tl	0.6	<12.5	<12.5	<12.5	<12.5	<12.5	<12.5	
	0.8	<12.5	<12.5	<12.5	<12.5	<12.5	<12.5	<12.5
	1	<12.5	<12.5	<12.5	<12.5	<12.5	<12.5	
V	0.6	203	243	223	247	295	271	
	0.8	280	329	305	265	300	283	124
	1	250	289	270	198	230	214	
Zn	0.6	71.6	<8.10	<39.9	77.8	<8.10	<43.0	
	0.8	(11.3)	<8.10	<9.7	19.9	<8.10	<14.0	38
	1	25.5	19.2	22.4	86	<8.10	<47.1	
Al[c]	0.6	6.21	5.99	6.1	0.761	0.792	0.777	
	0.8	5.73	6.48	6.11	2.7	2.69	2.7	2.17
	1	6.09	5.83	5.96	0.448	0.49	0.469	
Ca[c]	0.6	318	298	308	299	292	296	
	0.8	317	303	310	310	297	304	455
	1	302	296	299	306	303	305	
Fe	0.6	(47.7)	<37.8	<42.8	<37.8	<37.8	<37.8	
	0.8	<37.8	<37.8	<37.8	(39.1)	<37.8	<38.5	129
	1	97	<37.8	<67.4	56.4	<37.8	<47.1	
Mg	0.6	39.6	42.1	40.9	61.3	64.4	62.9	
	0.8	58.5	66.3	62.4	54.6	56.5	55.6	12,500
	1	38.4	34.5	36.5	43.1	36.8	40	

[a] These columns contain the concentrations for Replicate #1 and Replicate #2 of the indicated sample types.

[b] This column contains the concentrations for the extra sample from the Reactor B air heater outlet that was subjected to pH adjustment during the extraction process (see text).

[c] The concentration data for these metals are expressed in units of $\mu g/mL$ (ppm), rather than in the units indicated at the top of the column.

condensate on the surface of the ash, where it would be relatively easy to extract. But we currently have no explanation for the high extractability of Mo.

The data of Table 5-2 appear to suggest that exposure of the fly ash to ammonia enhanced the extractability of Se and possibly also Ba. But the apparent enhancement of Se extractability is probably due to the condensation of Se on the reactor outlet fly-ash particles, as explained above. The observed effect on Ba was quite weak; the average increase in Ba extractability due to ammonia exposure was 22%. This is such a small increment that it may not be real. In other words, a variation of this magnitude could have resulted from a combination of variability in sample composition and error in analytical measurement. However, the data do reflect a consistent upward trend in Ba extractability as the NH_3/NO_x ratio increases. This observation, together with the consistently higher Ba extractabilities in the reactor outlet samples (versus the inlet samples), leads us to suspect that the observed effect of ammonia on Ba extractability may be genuine, but leaves the unanswered question as to the mechanism for the effect.

None of the other detectable metals appeared to undergo an extractability enhancement due to ammonia exposure. Indeed, the ammonia exposure actually seemed to diminish the extractabilities of several metals, most notably Al, Cd, and Sb, and possibly also P, B, Rb, and As. Exposure to ammonia seemed to have no effect on the extractabilities of Cr, Mo, Co, Sr, V, Zn, Ca, and Mg.

Correlations between metal extractability and NH_3/NO_x ratio were frequently inconsistent and hence inconclusive. Apparently, the combined effects of variability in sample composition and variability due to analytical error were often sufficient to mask any effects caused by the intentional variation of the NH_3/NO_x ratio. As has already been noted, however, there was a consistent trend of increasing Ba extractability with increasing NH_3/NO_x ratio. In addition, the opposite trend, i.e., decreasing metal extractability with increasing NH_3/NO_x ratio, was observed for several other metals, e.g., Rb, Sb, As, and Mg. Note that, for each of these metals except Mg, this trend was consistent with the extractability-decreasing effect of ammonia exposure (versus no ammonia exposure) that was observed for these metals.

In the last column of Table 5-2, data are given for the sample in which a pH adjustment was attempted. In this column, one sees that the pH adjustment greatly enhanced the extractability of Mg from the fly ash. It may also have enhanced the extractabilities of Mn, Ca, As, and Fe. Some of these observations are explainable on the basis of the differing solubilities of the corresponding metal hydroxides at the two pH levels (see Table 5-1). For example, the solubility of $Mg(OH)_2$ at pH 11 is markedly lower than it is at pH 9, and a similar but smaller solubility difference exists for Mn. But this explanation may not fully account for all of the observed solubility increases.

QC Data

Our QC data generally indicated very few problems with interferences, calibration errors, and so on. Responses to both extraction blanks were uniformly below or (occasionally) very near our detection limits for all metals except Hg. For Hg, our blank responses were more than two-fold higher than our detection limit. But no Hg blank response above the detection limit was obtained for any of the real-sample extracts. Thus, the background Hg in the real-sample extracts may have been precipitated by the alkalinity or absorbed by the fly ash in these extracts.

Among the more revealing QC samples are the matrix spikes. Because they represent the recovery of spiked analyte from the actual sample-extract matrix, they are capable of identifying the presence of a variety of different types of matrix and spectral interferences. Hence, the matrix-spike recoveries are summarized in Table 5-3 and are discussed further here. In view of the very low spike levels that were used in this work, we considered the vast majority of these recoveries to be satisfactory for the intended purpose.

However, the recoveries for Sb and Fe were systematically high, which suggested a possible matrix or spectral interference in each case. Moreover, the recovery of Sn was erratic; this is a common problem with this element. But Sn and Fe were not detected at high enough levels in the sample extracts to enable us to draw any conclusions with regard to the effects of ammonia. Therefore, these interferences were of no particular significance to this study. But the problem with Sb is significant to the extent that the absolute magnitudes of the Sb found concentrations may be in error. However, the precision of the Sb responses was good, and thus the observed changes (i.e., trends) in Sb extractability as a function of ammonia exposure were, in our opinion, quite valid.

Conclusions

We concluded from this study that the extractability of Ba from fly ash by water is slightly enhanced by prior exposure of the fly ash to ammonia in the SCR system. In addition, the magnitude of this enhancement depends directly on the magnitude of the NH_3/NO_x ratio in the SCR unit. Of the 16 additional metals that could be detected in the fly-ash extracts, none displayed what we considered to be genuine enhancements in extractability, and several exhibited decreases in extractability as a result of exposure of the fly ash to ammonia. Although one of these metals -- Se -- displayed a large apparent increase in extractability on exposure to ammonia, we concluded that the Se found in the reactor-outlet sample extracts must have condensed from the gas phase onto the fly ash at the reactor outlet. Finally, a deliberate downward adjustment in the pH of one sample solution caused enhancements in the extractabilities of several metals, most notably Mg, but also Mn, Ca, As and Fe to a lesser degree.

Table 5-3. Recoveries of Matrix Spikes

	Matrix Spike Recovery, %				Matrix Spike Recovery, %		
Element	Matrix Spike #1	Matrix Spike #2	Average Recovery	Element	Matrix Spike #1	Matrix Spike #2	Average Recovery
Ag	96	104	100	P	110	118	114
As	91	108	100	Pb	122	124	123
B	—[a]	—[a]	—[a]	Rb	102	102	102
Ba	105	95	100	Sb	159	145	152
Be	87	86	87	Se	100	102	101
Cd	76	130	103	Sn	228	71	150
Ce	111	114	113	Sr	99	96	98
Co	84	94	89	Tl	106	108	107
Cr	108	118	113	V	100	100	100
Cu	121	128	125	Zn	91	94	93
Hg	89	93	91	Al	106	119	113
Mn	92	91	92	Ca	—[a]	—[a]	—[a]
Mo	104	100	102	Fe	127	193	160
Ni	90	102	96	Mg	104	89	97

[a] No matrix spikes were attempted for B and Ca because of their high natural concentrations in the samples (see text).

The results of our analyses raise several questions requiring discussion from theoretical perspectives:

* Should ammonia alter the extractability of any of the metals?

* Could the very marked apparent increase in the extractability of selenium actually be due to ammonia rather than the temperature change already suggested?

* How can the increase in extractability of barium be explained?

* How can the decreases in extractability of certain other metals be explained?

Increase in metal solubility due to complex ion formation with ammonia

Several metals react with ammonia to form a series of complex ions. This phenomenon is illustrated by the following reactions beginning with Co^{+2} ion:

$$Co^{+2} + NH_3 = Co(NH_3)^{+2} \qquad K_1$$

$$Co(NH_3)^{+2} + NH_3 = Co(NH_3)_2^{+2} \qquad K_2$$

$$Co(NH_3)^{+2} + NH_3 = Co(NH_3)_3^{+2} \qquad K_3$$

$$Co(NH_3)^{+2} + NH_3 = Co(NH_3)_4^{+2} \qquad K_4$$

The enhanced solubility of cobalt can be expressed as the ratio R of total dissolved cobalt, including the complexed and uncomplexed metal, to the uncomplexed metal as a function of the several complexation constants and the concentration of excess, uncomplexed ammonia:

$$R = [Total\ Co]/[Uncomplexed\ Co]$$

$$= 1 + K_1[NH_3] + K_1 K_2[NH_3]^2 + K_1 K_2 K_3[NH_3]^3 + K_1 K_2 K_3 K_4[NH_3]^4$$

The values of the logarithms of these complexation constants involving Co^{+2} and NH_3 are approximately as follows:

$$\log K_1 = 1.99$$

$$\log K_2 = 3.50$$

$$\log K_3 = 4.43$$

$$\log K_4 = 5.07$$

If the concentration of uncomplexed NH_3 is, for example, 0.009 M, the value of the ratio R is 1.32. If there were no other form of cobalt in solution, then, the presence of NH_3 in an amount providing an uncomplexed concentration of 0.009 M would increase the solubility of cobalt by 32%.

The compilation of complexation constants published by Bjerrum indicates that there are six metal ions that are significantly complexed by ammonia. These are the ions of cadmium, cobalt, copper, nickel, silver, and zinc. All but silver produce at least four complexes as illustrated above for cobalt; silver, on the other hand, produces only the complexes with one and two moles of ammonia. Bjerrum lists several sets of values for the complexation constants of each metal; the reader cannot easily establish the best set of values for each metal. For the present purposes, however, a set of approximately correct values has been selected for each metal; these sets of values are presented in Table 5-1 (the selected set for cobalt is the same as the one listed above in the text). The ratios of concentrations of each metal in all forms to that in uncomplexed form have subsequently been calculated, with the results given in Table 5-2. As before, the uncomplexed concentration of NH_3 was assumed to be 0.009 M (a value to be commented upon momentarily).

The concentrations ratios listed in Table 5-2 are provisional values only, which require further discussion. They indicate, however, that ammonia has a far greater solubilizing effect on copper than on any of the other metals, because of course the complexes based on copper are more stable than those of any other metal. For copper, the multiplying factor is 5430; for cobalt it is just 1.32.

The assumed value of 0.009 M for uncomplexed ammonia is more than of incidental value. Suppose that ammonia is retained by fly ash at a concentration of 1000 μg/g, a value just slightly higher than that measured for any NH_3/NO_x ratio in this investigation. If this ash is then placed in water in the ratio of 5 g per 100 mL and all the ammonia goes into solution, the ammonia concentration will be 0.009 M. Moreover, if only a very much smaller concentration of a metal is extracted, as in all solutions analyzed in this investigation, the concentration of uncomplexed ammonia will remain near 0.009 M. This concentration, then, and the results in Table 5-2 based on this concentration have a provisional applicability for predicting enhanced metal extractabilities in this investigation.

None of the observed enhancements that can be assigned numerical values clearly conforms to the prediction. This perhaps can be said most emphatically for copper, for which the enhancement seems certainly not to be three orders of magnitude. Whether silver satisfies the prediction of an enhancement by the factor 158 cannot be said, because this metal was not measurable with or with ammonia added. Cadmium is not enhanced by

J. Bjerrum et al., Stability Constants of Metal-Ion Complexes, with Solubility Products of Inorganic Substances, Part II: Inorganic Ligands, The Chemical Society, London, 1958.

a factor of 2.26 but instead seems to have been diminished significantly. Cobalt, nickel, and zinc are difficult to discuss because of the ambiguous data.

What can be said about the erroneous prediction for copper? Probably the explanation lies in the amphoteric character of this metal, which makes the simple uncomplexed metal ion relatively unimportant at the high pH values of the ash extracts investigated. The amphoteric character of copper is illustrated by the following equation:

$$Cu^{+2} + 4OH^- = Cu(OH)_4^{-2}$$
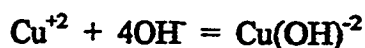
The equilibrium constant for this process is about 1×10^{16} and at pH 11 the concentration ratio of the hydroxide complex (more often called cuprate ion) to the uncomplexed ion is about 10,000. Thus, the complexing action of hydroxide is greater than that of ammonia, and the large enhancement in solubility due to ammonia cannot be readily seen. If ammonia increases the solubility of the free metal ion by the factor 5,000 and hydroxide does so by the factor 10,000 at pH 11, the net effect of ammonia alone will be to increase total solubility by just the factor 1.5.

Zinc also is amphoteric and would have the solubilizing effect of ammonia damped by the corresponding but stronger effect of hydroxide. The other four metals of concern are not amphoteric but still, with the exception of silver, would not be strongly affected by ammonia. As indicated before, silver may be influenced strongly by ammonia in a relative sense but not enough in an absolute sense for the effect to be measured.

Increase in the extractability of selenium

There is the possibility of a reaction between SeO_2, the volatile form of selenium, and NH_3 to cause deposition of selenium on the ash surface. The reaction is the reduction of the oxide to the much volatile metal:

$$3SeO_2 + 4NH_3 = 3Se + 2N_2 + 6H_2O$$
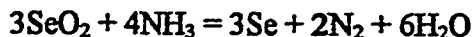
The literature discusses this reaction and cites the utility of the reaction for refining elemental selenium from the oxide (the process entailing oxidation of the crude element to the volatile oxide and the subsequent reduction of the latter with NH_3). Chemists at Southern Research Institute have identified the reaction as an explanation for deposition of an orange film (elemental selenium) on baghouse filters when ammonia is used for conditioning the filter cake to lower the pressure drop across the deposit.

We cannot state which is the more probable process -- deposition of SeO_2 as the result of the temperature reduction or deposition of elemental selenium as the result of the oxide reduction with ammonia. Attributing the increased extraction of selenium to the first of these processes seems the more natural choice, however, because of the very sharp reduction in volatility of

SeO_2 between 700° and 300° F. The reduction in volatility from that of SeO_2 to that of the element at 300° F provides a less dramatic driving force for the deposition of selenium.

Increase in extractability of barium

Bjerrum gives no indication that barium reacts with ammonia to produce complex ions. Thus, if ammonia is the cause of increased extractability of this metal, the effect must be through another mechanism, none of which is apparent as a direct process. Conceivably, the effect on barium is not due to ammonia even indirectly. It may be the result of the change in temperature that accompanied the addition of ammonia (because samples containing ammonia were collected after the temperature had decreased in a heat exchange process). No way to explain an increased extractablity of barium at a lower sampling temperature, however, is evident. Specifically, no process analogous to the condensation of SeO_2 on ash surfaces as the temperature falls can be postulated. Barium is essentially as involatile at the higher temperature (around 700° F) as at the lower temperature (300° F).

Decrease in extractability of other metals

These effects occurred with several metals and require comment just as does the contrary effect with barium. It is reasonable to ask the question of whether processes can be suggested whereby ammonia directly or indirectly lowers metal solubility. There are rare processes if any that produce relatively insoluble aggregates of a metal ion with ammonia. None can be suggested that are associated with those metals that seems to have decreased extractability from fly ash as the result of the presence of ammonia.

Aluminum is perhaps the metal with the greatest apparent reduction in extractability with ammonia present. This metal has no reaction with ammonia known to the present investigators. Its behavior is not readily explained by the decrease in temperature that accompanied the addition of ammonia, nor is it explained by the other phenomenon known to occur — the increased pH of the extract. In fact, aluminum is amphoteric and if influenced by the change in pH it should have increased in extractability, not decreased.

INNOVATIVE CLEAN COAL TECHNOLOGY (ICCT)

Demonstration of Selective Catalytic Reduction (SCR)
Technology for the Control of Nitrogen Oxide (NO$_x$)
Emission from High-Sulfur, Coal-Fired Boilers

Economic Evaluation
of Commercial-Scale SCR Applications
for Utility Boilers

September 1996

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CONTENTS

LISTING OF FIGURES

LISTING OF TABLES

LISTING OF EXHIBITS

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ACKNOWLEDGEMENTS

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EXECUTIVE SUMMARY

This report presents the results of an economic evaluation produced as part of the Innovative Clean Coal Technology project, which demonstrated selective catalytic reduction (SCR) technology for reduction of NO_x emissions from utility boilers burning U.S. high-sulfur coal. The project was sponsored by the U.S. Department of Energy (DOE), managed and cofunded by Southern Company Services, Inc., (SCS) on behalf of The Southern Company, and also cofunded by the Electric Power Research Institute (EPRI) and Ontario Hydro.

The document includes a commercial-scale capital and O&M cost evaluation of SCR technology applied to a new facility, coal-fired boiler utilizing high-sulfur, U.S. coal. The base case presented herein determines the total capital requirement, fixed and variable operating costs, and levelized costs for a new 250-MW pulverized coal utility boiler operating with a 60-percent NO_x removal. Sensitivity evaluations are included to demonstrate the variation in cost due to changes in process variables and assumptions.

This report also presents the results of a study completed by SCS to determine the cost and technical feasibility of retrofitting SCR technology to selected coal-fired generating units within the Southern electric system. While retrofit issues will vary from plant to plant and company to company, the results of this study reflect the typically wide range of retrofit costs due to site-specific issues encountered at those plants studied.

The conclusion shows the 250-MW base case unit capital and first year O&M (in 1996 dollars) are $13,415,000 ($54/kW) and $1,045,000, respectively. Levelized cost for the base case unit is $2,500/ton on a current dollar basis and $1,802/ton on a constant dollar basis. Busbar cost is 2.57 mills/kWh on a current dollar basis and 1.85 mills/kWh on a constant dollar basis.

For the new plant applications, total capital requirement for a 60 percent NO_x removal design ranged from $45/kW for a 700-MW unit to $61/kW for a 125-MW unit. Associated current dollar levelized cost ranged from $2,165/ton to $2,811/ton for the 700-MW unit and 125-MW unit, respectively.

Capital cost variation as a function of NO_x removal for a 250-MW unit ranged from $57/kW for an 80 percent design to $52/kW for a 40 percent removal design. Corresponding current dollar levelized cost ranged from $2,036/ton to $3,502/ton for the 80 percent and 40 percent removal cases, respectively.

Retrofit applications for a 60 percent removal design show a range of capital requirements from $59/kW for an 880-MW unit size to $87/kW for a 100-MW units size. There are two plants having capital requirements of $130/kW and $112/kW due to balanced draft conversion of the units. Levelized costs range from $1,848/ton to $5,108/ton on a current dollar basis.

Demonstration of Selective Catalytic Reduction (SCR) Technology for the Control of Nitrogen Oxide (NO$_x$) Emission from High-Sulfur, Coal-Fired Boilers
DOE ICCT Project DE-FC22-90PC89652

Economic Evaluation
of Commercial-Scale SCR Applications for Utility Boilers

1.0   Introduction

This report presents the results of an economic evaluation produced as part of the Innovative Clean Coal Technology project, which demonstrated selective catalytic reduction (SCR) technology for reduction of NO$_X$ emissions from utility boilers burning U.S. high-sulfur coal. The project was sponsored by the U.S. Department of Energy (DOE), managed and cofunded by Southern Company Services, Inc., (SCS) on behalf of The Southern Company, and also cofunded by the Electric Power Research Institute (EPRI) and Ontario Hydro. Six world-wide catalyst suppliers and major equipment suppliers also participated with technical and financial contributions to the project. The project was located at Gulf Power Company's Plant Crist Unit 5 (75-MW tangentially fired boiler) located near Pensacola, Florida. The test program was conducted for approximately 2 years to evaluate catalyst deactivation and to quantify operational impacts of SCR technology employed in a high-sulfur environment. The SCR test facility included nine reactors: three 2.5-MW (large) reactors rated at approximately 5000 scfm (8500 Nm^3/hr) and six 0.2-MW (small) reactors rated at approximately 400 scfm (680 Nm^3/hr). Eight reactors operated in a hot-side, high-dust configuration while the ninth reactor operated in a hot-side, low-dust configuration. All reactors operated in parallel with commercially available SCR catalysts.

Ultimately, the goal of any test facility is to gather information and gain experience to enable a more accurate performance evaluation as well as economic analysis when extrapolated to commercial size installations. From its inception, the SCS/DOE test facility was designed to minimize the uncertainty associated with application of pilot scale test results to a full scale installation. Significant resources have been expended to present realistic costs and performance expectations of SCR technology based on the results of the 2-year test program. It is anticipated that the economic analysis presented in this report will assist interested parties with evaluating SCR compared to other possible NO$_X$ control alternatives for future emission control requirements.

There are several regulatory and environmental drivers in various stages of consideration which may increase the likelihood of employing SCR technology in the future. Recent experience of applying SCR to new coal-fired installations has created regulatory precedent under New Source Review, which will affect future best available control technology (BACT) and lowest achievable emission rate (LAER) determinations for other new units. With one exception, these new installations are owned and/or operated by independent power producers (IPPs) who report that adopting SCR technology was necessary to quickly obtain the construction and/or operating permits.

The 1990 Clean Air Act Amendments (CAAA) mandated several NO_x control requirements and regulatory reviews to reduce NO_x emissions from utility boilers. Application of SCR to existing boilers is being considered for units located in areas designated under Title I (nonattainment provisions) for attainment of the ambient ozone standard. Recent efforts by the Ozone Transport Assessment Group (OTAG) have focused on NO_x reduction strategies on a broader scale, encompassing all states in the central and eastern part of the United States. Results of the OTAG review may increase the likelihood for retrofits of SCR technology, particularly if emission averaging and NO_x trading are allowed. Additionally, nationwide reductions in NO_x mandated under Title IV (acid rain provisions) will be required by the year 2000. In order to meet these additional NO_x reductions, utilities are given flexibility in selecting the most suitable and cost-effective NO_x control technologies for their situation.

This report is written from the perspective of a utility end user of SCR technology. As such, the results are meant to establish a range of financial exposure representative of most domestic electric utilities. It is recognized that there will be utility specific instances where the cost (or cost effectiveness) of SCR technology may be higher or lower than what is contained in this report as evidenced in previous papers representing diverse views regarding the cost of SCR technology. (refer to section 5.0 for a list of reference papers). In an effort to present the most effective economic evaluation possible, information was obtained and incorporated from several sources including:

- Test Facility Data - Measured data and operational lessons learned at the SCS/DOE test facility over the 2-year test program formed the basis of the technical performance estimates.

- Peer Review - Comments were solicited from cofunders, project participants, and independent consultants. The review cycle accomplished a key objective of obtaining peer review of the material as well as challenging the results based on differing viewpoints.

- Technology Suppliers - Analytical and engineering analysis received from vendor participants contributed greatly to the success of the project. The catalyst management plans presented in this report are based on vendor generated laboratory data of catalyst deactivation (k/ko) over time. Additionally, air preheater performance, material testing, and deposit analysis were supplied by the air preheater vendor.

- Full-Scale, Coal-Fired Experience - Results of the economic analysis are enhanced by incorporating current market trends based on SCS participation in one of the first commercial coal-fired SCR installations in the United States. Information from the other U.S. coal-fired SCR installations was also considered when developing the economic evaluation.

The economic evaluation presented in this report is not meant to supplant the need to perform site-specific financial and pro-forma analyses when evaluating SCR technology for a specific project. It is recognized that there will likely be project-specific constraints, sensitivity analyses, and market forces which no generalized economic evaluation will capture. Rather, the

information reported herein is presented so the user can modify key financial and technical assumptions to customize the results to a specific situation.

Section 1.0 of this document provides a brief overview of the project and outlines major market drivers for consideration of SCR technology for future NO_x reduction requirements.

Section 2.0 presents a commercial-scale capital and O&M cost evaluation of SCR technology applied to a new facility, coal-fired boiler utilizing high-sulfur, U.S. coal. The base case presented herein determines the total capital requirement, fixed and variable operating costs, and levelized costs for a new 250 MW pulverized coal utility boiler. Economic factors are calculated according to guidelines established by EPRI, taking into account financial parameters such as the cost of capital, income tax rates, and the rate of inflation. Two different sets of factors are calculated to permit the economics to be presented either on a current dollar basis, which includes the effect of inflation, or constant dollar basis which ignores inflation. Reporting of the results are based on "General Guidelines for Public Design Report and Final Report" prepared by Burns and Roe Services Corporation for the DOE Pittsburgh Energy Technology Center (PETC).

Section 3.0 contains sensitivity evaluations which are included to demonstrate the variation in cost due to changes in process variables and assumptions. The following sensitivity cases are included in this evaluation:

- Capital, O&M, and levelized cost for new SCR vs. unit size (60 percent NO_x removal).

- Capital, O&M, and levelized cost for new SCR vs. NO_x removal efficiency (250-MW plant size).

- Levelized cost for new SCR vs. inlet NO_x concentration (250-MW plant size and 60 percent NO_x removal).

- Levelized cost for new SCR vs. catalyst relative activity (catalyst management plans for 250-MW plant size and 60 percent NO_x removal).

- Levelized cost for new SCR vs. return on common equity (ROE for 250-MW plant size and 60 percent NO_x removal).

- Capital, O&M, and levelized cost for new SCR vs. catalyst price (250-MW plant and 60 percent NO_x removal).

Section 4.0 presents the results of a study completed by SCS to determine the cost and technical feasibility of retrofitting SCR technology to selected coal-fired generating units within the Southern electric system. While not the direct result of the SCS/DOE test facility, many of the same methodologies and lessons learned have been applied to utility-scale applications in an effort to maximize the value of the test facility investment to The Southern Company. While retrofit issues will vary from plant to plant and company to company, the results of this study reflect the typically wide range of retrofit costs due to site-specific issues encountered at those plants studied within the Southern electric system.

Section 5.0 contains a list of references which were consulted for supplemental information included in this document.

3

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2.0 Application of SCR Technology For a New Unit

2.1 Technical Premises

The economic evaluation presented in this section is based on the application of a high-dust, hot-side SCR configuration (i.e., located between the boiler economizer outlet and the air preheater inlet) at a new coal-fired facility. Where applicable, design premises that have a major impact on cost estimating are described in more detail.

The technical design premises used to prepare the economic analysis were selected to be representative of actual or anticipated plant configurations and NO_x control requirements currently being permitted or likely to be permitted on new coal-fired boilers in the United States. Therefore, defining assumptions were selected in an effort to have broad utility applicability. The following paragraphs describe major features of the base case installation.

2.2 250-MW Base Case Unit Description

The base case represents a new, base-load 250-MW pulverized-coal power plant typical of the majority of new coal-fired projects currently under development, construction, or recently declared in commercial operation. The 250 MW plant size is consistent with current and future capacity trends of new domestic power plants. The plant is located in a rural area with minimal space limitations. The fuel is a high-sulfur bituminous Illinois No. 6 coal having an analysis shown in table 1.

Table 1
Coal Analysis Used for Economic Evaluation

Proximate Analysis	Dry Basis	As Received
Ash	9.30 %	8.39 %
Volatile Matter	37.88 %	34.16 %
Fixed Carbon	52.82 %	47.65 %
Moisture		9.80 %
Total	100.00 %	100.00 %

Ultimate Analysis	Dry Basis	As Received
Carbon	74.82 %	67.48 %
Hydrogen	5.00 %	4.51 %
Nitrogen	1.58 %	1.43 %
Sulfur	2.58 %	2.33 %
Chloride	0.16 %	0.14 %
Oxygen	6.56 %	5.92 %
Ash	9.30 %	8.39 %
Water		9.80 %
Total	100.00 %	100.00 %

Higher Heating Value (HHV)	13,265 Btu/lb	12,500 Btu/lb

The plant utilizes a single, balanced-draft, pulverized-coal fired boiler complete with all required auxiliary equipment. The boiler is designed to produce approximately 1,610,000 lb/hr of main steam at turbine inlet conditions of 2400 psig and 1000°F. Utilizing current generation low-NO_x combustion systems, the boiler will produce a NO_x emission rate of 0.35 lb/MBtu. For purposes of this study, it is assumed that tangentially fired boilers and wall-fired boilers are interchangeable with respect to thermal performance and flue gas constituents.

Coal is delivered through gravimetric feeders to the pulverizers and then to the coal nozzles located in the furnace walls. Primary combustion air will flow through the pulverizers to transport the pulverized coal to the furnace. Secondary combustion air from the forced draft fans is preheated in the air preheater and will then be ducted to the boiler windbox to be injected into the furnace through the burners and overfire air ports.

The flue gas exits the boiler and enters a single, hot-side SCR reactor. Flue gas flow is vertically downward through the reactor. The physical arrangement of the SCR is located directly above the air preheater. The SCR is designed as a universal reactor able to accept either (or both) plate- or honeycomb-type catalysts. Nominal generic catalyst module dimensions of 2 meter (l) x 1 meter (w) x 1 meter (h) were assumed for this study. Anhydrous ammonia is used as the reagent. Ammonia injection dilution air will utilize stand alone air fans rather than combustion air from the primary air system.

A single, trisector, Ljungstrom regenerative air preheater is utilized to reclaim heat from the flue gas stream and transfer that heat to the primary and secondary air. The heat transfer surface arrangement includes hot, intermediate, and cold sections. Physical features of the air preheater are typical of what is commercially offered as a deNO$_x$ air preheater as mentioned later in this report. As a result of the air preheater materials testing, the intermediate and cold end heat transfer surface are enamel coated.

Sulfur dioxide removal is accomplished by a lime spray dryer flue gas desulfurization (FGD) system. The FGD system includes two 50-percent absorber vessels equipped with rotary atomizers that produce very fine droplets to enhance the reactivity of the slurry. The absorber vessels will be designed with sufficient residence time to ensure complete evaporation of the water and collection of the acid gases.

A reverse gas, fabric-filter baghouse is used to collect the dried reaction products from the spray dryer as well as the flyash produced in the boiler by the combustion of coal. The baghouse will be constructed in multiple compartments that allow on-line cleaning and maintenance. Each compartment is equipped with a single ash hopper. Clean gas from each compartment passes to an outlet manifold common to all compartments. The clean gas exits out of the baghouse and to the induced draft (ID) fans for discharge out the stack.

Assumptions used to prepare the material balance and combustion calculations for the 250-MW base case unit are shown in table 2. The combustion calculation output for the 250-MW base case is shown in exhibit B.

6

Table 2
250-MW Base Case Material Balance and Combustion Calculation Assumptions

Unit Capacity (Gross)	250 MW
Capacity Factor	65%
Type of Installation	New facility
Boiler Type	Wall-fired or tangentially fired
Heat Input	2375 MBtu/hr
Coal Feed	190,000 lb/hr
Gross Plant Heat Rate	9500 Btu/kWh
Type of Air Preheaters	Vertical shaft, Ljungstrom
Number of Air Preheaters	One
Air Preheater Outlet Temperature	300°F
Air Preheater Leakage	13%
Excess Air @ Boiler Outlet	18%

2.3   250-MW Base Case SCR Design Criteria

General design criteria for the SCR assumed for this study are shown in table 3. This criteria is predominantly based on the design of the SCR test facility as previously reported in "Plant Crist SCR Project SCR Test Facility Design Basis," Volume 1 and 2 submitted to DOE as the Public Design Report. Where applicable, design criteria have been modified to better reflect operational lessons learned from the test facility and/or current utility industry trends in post combustion NO_x control.

Table 3
250-MW Base Case SCR Design Criteria

Type of SCR	Hot-side
Number of SCR Reactors	One
Reactor Configuration	3 catalyst support layers + 1 dummy layer
Initial Catalyst Load	2 of 3 layers loaded, 1 spare layer
Required Range of Operation	35% to 100% boiler load
NO_x Concentration @ SCR Inlet	0.35 lb/MBtu
Design NO_x Reduction	60%
Flue Gas Temp @ SCR Inlet	700°F
Flue Gas Pressure @ SCR Inlet	-5 in. W.G.
Design Ammonia Slip	5 ppm
Guaranteed Catalyst Life	2 years (16,000 hours)
SO_2 to SO_3 Oxidation	0.75% (initial catalyst load)
Maximum Pressure Drop	6 in. W.G. (fully loaded reactor)
Velocity Distribution	$\Delta V / V_{mean} < 10\%$ over 90% of reactor area
	$\Delta V / V_{mean} < 20\%$ over remaining 10% area
Ammonia Distribution	$\Delta C / C_{mean} < 10\%$
Temperature Distribution	$\Delta T < 10°C$ max deviation from mean

7

Specific design criteria and technical assumptions which have a major impact on capital and operating cost estimation are described in more detail in the following paragraphs.

2.3.1 SCR Reactor

The following assumptions were used in the development of the SCR reactor capital cost:

- Although the test facility reactors were designed with four catalyst layers plus one flow straightener (dummy layer), this configuration is not thought to be representative of current commercial trends for new units equipped with state of the art low-NO_x burner technology. The test facility reactors were designed with maximum possible flexibility in anticipation of potential problems developing during the test program. The spare layer was not utilized by any catalyst vendor during the test program. It was also assumed that for a new facility with no space limitations, the cross section of the reactor and the height of the catalyst modules could be adjusted within acceptable ranges to allow the initial load of catalyst to be housed in two layers rather than three. Thus, the reactor assumed for this study utilizes a configuration with three catalyst layers plus a flow straightener layer.

- A single, vertical downflow reactor is provided.

- The flow straightener layer consists of fabricated modules of 2 in. x 2 in., 16-gauge mild steel tube approximately 18 inches in length. The design objective for the flow straightener is to ensure that the ratio of hydraulic diameter of the channel openings to the length of flow is sufficient to produce vertical flow streamlines at the inlet to the first layer of catalyst.

- The reactor is equipped with an economizer bypass to permit SCR operation at lower boiler loads. The economizer bypass was sized to allow up to 5 percent of the boiler flue gas flow. It is recognized that the economizer bypass may be different in size or eliminated completely depending on project specific requirements.

- Consistent with many of the new commercial installations, the SCR reactor was assumed to be integral to the boiler house structure and enclosed with a roof and siding.

- All catalyst layers include steam sootblowers. The sootblower design is identical to those used in the test facility.

2.3.2 Initial Space Velocity and Catalyst Volume

Space velocity is a process variable which is used in determining the quantity of catalyst required for a given NO_x removal requirement. Space velocity is defined as the volume of flue gas treated per unit volume of catalyst. The standard convention for expressing flue gas flow rate is in ft^3/hr (m^3/h) corrected to conditions of 32°F (0°C) and 1 atmosphere (1 bar). Catalyst volume is expressed in corresponding units of ft^3 or m^3. Thus, space velocity can be expressed:

$$SV \ (1/hr) \ = \ \text{Flue Gas Flow } (ft^3/hr \text{ or } m^3/hr) \ / \ \text{Catalyst Volume } (ft^3 \text{ or } m^3)$$

The relationship between initial space velocity and NO_x removal used in this evaluation is shown in figure 1. The relationship for new units is represented by a least squares curve fit of space velocities taken from the five new coal-fired SCR installations in the U.S. Design information was assembled from commercial bid evaluations, project specific design criteria, and publicly available technical literature. A total of nine data points indicative of both honeycomb- and plate-type catalysts was used to construct the curve. Thus, depending on the project specific evaluation and catalyst geometry selected, the actual space velocity may be slightly higher (as in the case of honeycomb catalyst) or slightly lower (in the case of plate catalyst) than the indicated curve.

The relationship for retrofit units was developed using a least squares curve fit of test facility data measured during parametric testing and each catalyst supplier's proposed space velocity based on the test facility steady state design criteria. The relationship does not represent a single catalyst supplier's offering, but rather a composite of all catalyst space velocities. This approach was selected to provide a reasonable method for estimating space velocity which is independent of catalyst geometry. From a user perspective, this permits consideration of a reactor capable of housing different catalysts ("universal reactor") to allow end-users to evaluate different catalyst offerings directly from the catalyst supplier rather than through a process or system supplier.

2.3.3 Catalyst Life and Catalyst Management Plan

The term "catalyst life guarantee" is often misinterpreted to mean that the performance of the SCR sharply decreases and the entire volume of catalyst must be replaced after the guarantee period. This interpretation is not correct. Performance during early project years (or months) normally exceeds the guaranteed values. Over time, the catalyst performance will gradually deteriorate until the SCR is unable to maintain the required NO_x removal while simultaneously achieving the required ammonia slip. (Most SCR installations operate on a constant NO_x removal to allow continued operation with permitted NO_x emissions at the expense of increased ammonia slip.) Even though the SCR cannot meet guaranteed ammonia slip vs. NO_x performance, the catalyst still has considerable activity remaining.

As noted above, the SCR reactor for this evaluation includes space for three catalyst layers plus a flow straightener. At time zero, two of the three catalyst layers are loaded with catalyst. The third layer, which is empty, allows catalyst suppliers to develop optimized catalyst management plans which increase catalyst utilization. Thus, a fresh catalyst layer can be added to the reactor after the guarantee period when the ammonia slip begins to exceed the guaranteed limit. The activity of the new catalyst combined with the residual activity of the existing catalyst restores the performance of the SCR and extends the next addition/replacement outage beyond the initial guarantee interval.

Catalyst deactivation data were periodically measured by taking catalyst samples from the test facility reactors and returning the samples to the respective catalyst supplier. The catalyst suppliers performed a standard protocol of laboratory and bench scale tests to develop an activity

Space Velocity vs. NOx Removal
SCS/DOE Project – Economic Evaluation

New Unit: SV = 8020 x e (-7.77489E-3 x [%NOx Red])
(Carney's Point, Logan, Indiantown, Birchwood, Stanton)

Retrofit Unit: SV = 5506 x e (-6.95843E-3 x [%NOx Red])
(SCS/DOE Test Facility Design Information)

NOx Removal (%)

Figure 1

Space Velocity (1/hr)

DOESV.atb

10
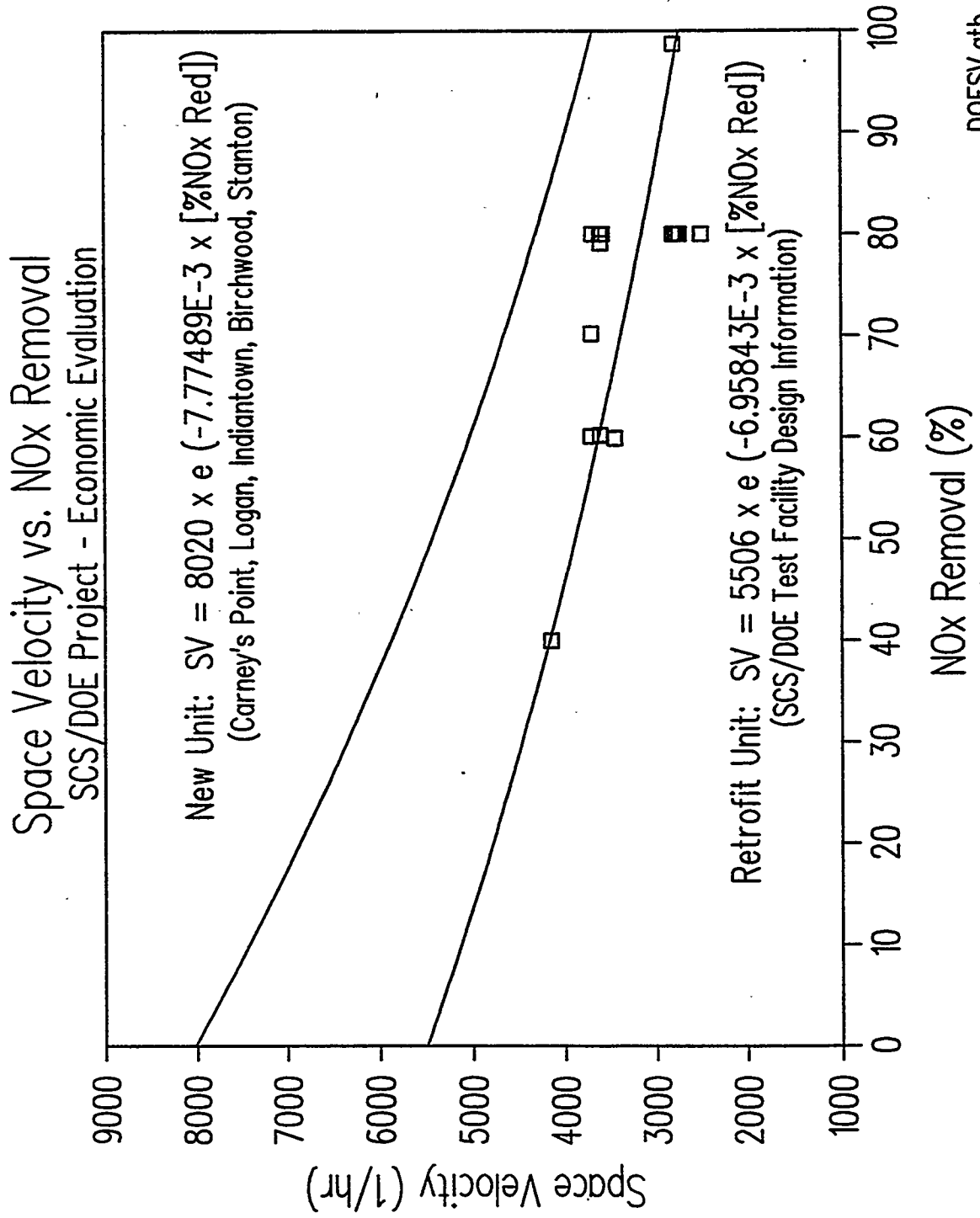
decline vs. operating time relationship. The base case catalyst management plan shown in figure 2 was derived using data collected at the test facility from all catalyst suppliers. Figure 3 shows that a least squares curve fit of catalyst relative activity data from the test facility results in a k/ko value of approximately 0.80 after 16,000 hours. Refer to the catalyst management plan sensitivity section of this report for additional discussion regarding volatility of the k/ko data.

The management plan is based on a 16,000-hour (2-year) catalyst life guarantee period. After the initial guarantee period of 2 years, a new layer of catalyst is added to the reactor spare layer, thus taking advantage of the residual activity in the initial layers to boost the performance of the SCR. The next addition of catalyst is required in project year 6, when one of the initial layers is replaced. After year 6, staged replacement of catalyst layers occurs approximately every three years over the remaining life of the project.

Because the majority of SCR installations contractually obligate the catalyst (or process) supplier to dispose of spent catalyst as part of the initial contract, catalyst disposal costs are not included as part of these cost estimates. This obligation typically is not contingent on catalyst replacement sales. The user pays all shipping costs for transporting the spent catalyst back to the supplier where it is recycled and/or reclaimed. One catalyst supplier has identified a party interested in reclaiming the vanadium as a feedstock for other industrial uses.

2.3.4 Air Preheater

The incremental cost of a deNO$_x$ air preheater (APH) over a non-SCR application air preheater is included in the economic evaluation. Further, based on test facility results provided by ABB Air Preheater, Inc., their recommendation of utilizing enamel coating for the intermediate and cold end heat transfer surface is also included in the evaluation. The following summarizes the air preheater assumptions:

- A single Ljungstrom, regenerative trisector air preheater.

- Intermediate heat transfer surface constructed of 20-gauge (U.S.) low-alloy, corrosion-resistant material. Cold end heat transfer surface constructed of 18 gauge (U.S.) low-alloy, corrosion-resistant material. Normal construction without SCR is 24-gauge (U.S.) open hearth material.

- Intermediate heat transfer surface fabricated with notched flat, 6mm (NF6) surface profile. Normal intermediate surface profile is a more efficient double undulating (DU) surface profile. Cold end heat transfer surface fabricated with NF6 surface profile. Normal cold end surface profile is more efficient with the NF3.5 surface profile.

- With looser, less efficient heat transfer surface in the intermediate and cold end sections, more heat transfer surface will be required to maintain a net zero impact on the thermal performance of the air preheater. More surface area translates to larger, heavier air preheater housing and rotor which requires an upgrade in the support bearing.

11

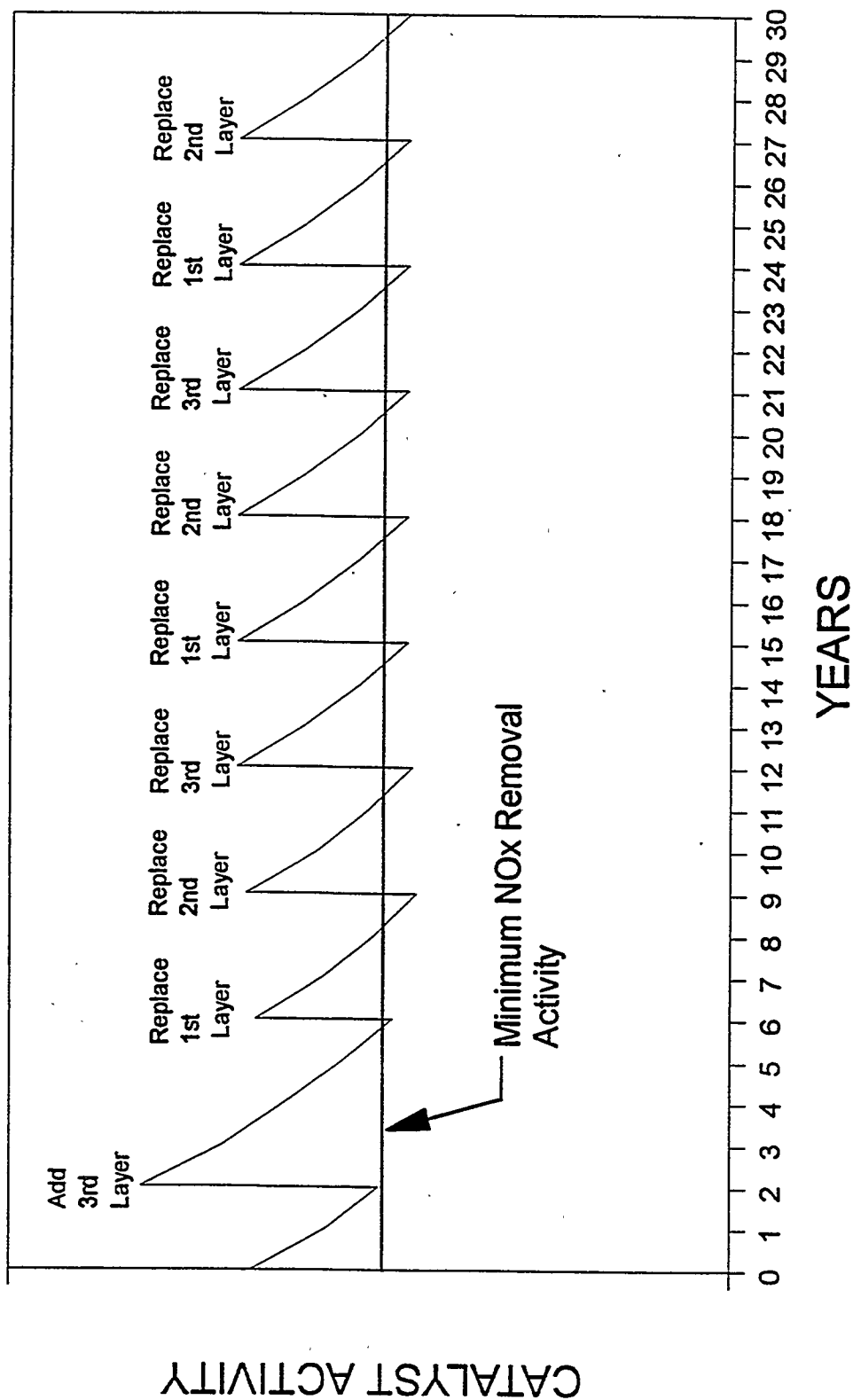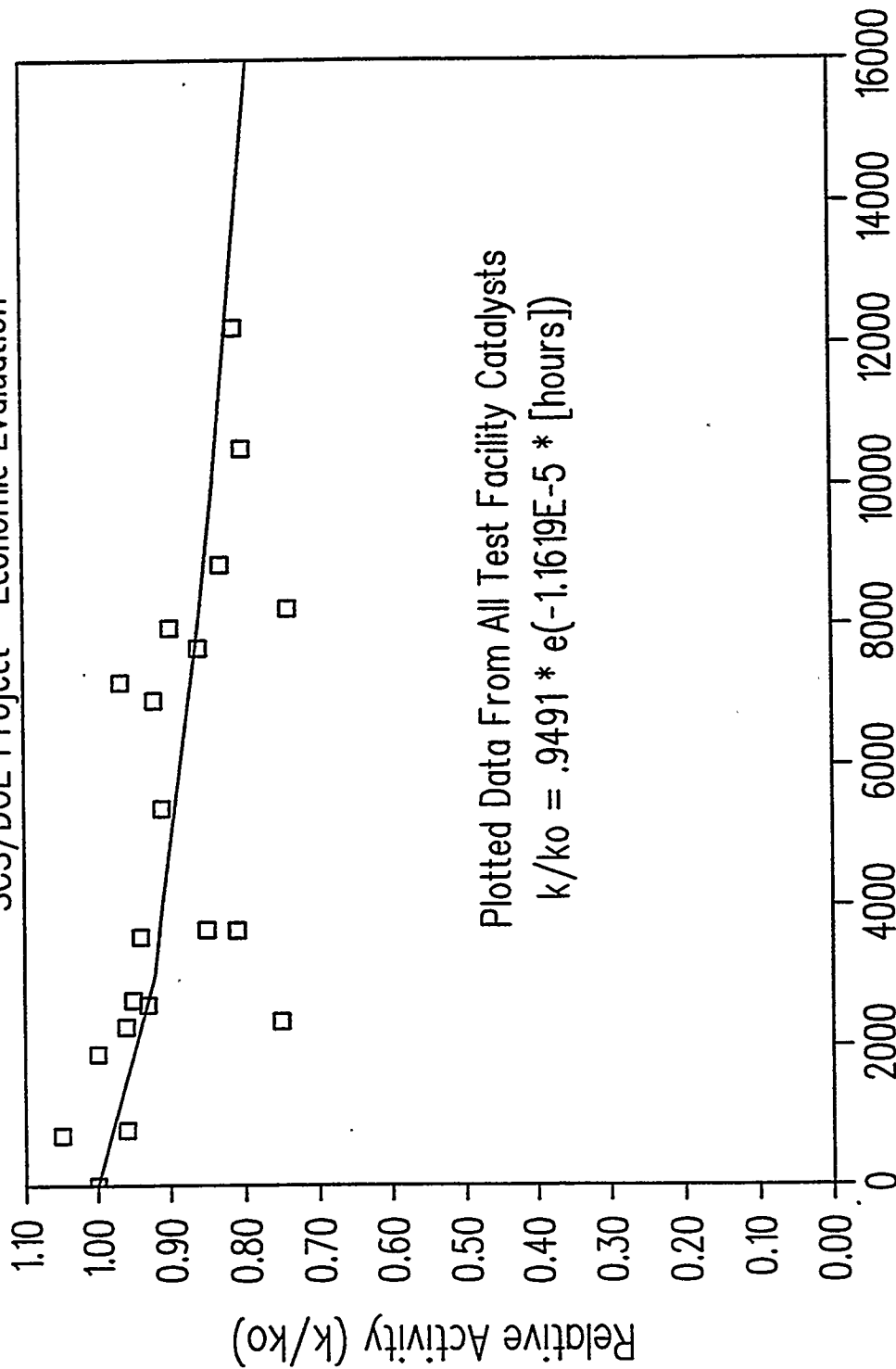CATALYST MANAGEMENT PLAN
DOE SCR Project - Economic Evaluation

Figure 2

Relative Activity Data (k/ko) vs. Time
SCS/DOE Project - Economic Evaluation

Plotted Data From All Test Facility Catalysts

$$k/ko = .9491 * e(-1.1619E-5 * [hours])$$

Relative Activity (k/ko)

Operating Time (hours)
Figure 3

DOEK_Ko1.atb

13.

- As a result of a heat transfer material evaluation performed at the test facility, ABB Air Preheater, Inc., has recommended the use of enamel coating on the intermediate and cold end heat transfer surface. Therefore, this incremental cost is included in the estimate. This also will have an impact on the air preheater weight, requiring further upgrade in support bearings.

- Additional steam sootblowers and water washing equipment are provided on both hot and cold ends of the deNO$_x$ air preheater.

- Slightly higher air leakage rates can be expected with incrementally lower flue gas static pressures.

2.3.5 ID Fan

Comparison of ID fan duty with and without SCR will indicate a differential cost due to increased flow and static pressure requirements for the ID fan with the SCR installation. The economic evaluation includes the incremental capital and O&M cost for the ID fan. The following summarizes the ID fan assumptions:

- Two 55-percent capacity fans with direct drive electric motors and inlet vane control.

- It was assumed that the SCR would increase the test block static pressure +9 inches W.G. This corresponds to a system pressure drop of approximately +6 inches W.G. with a fully loaded reactor. A 2-percent increase in flue gas mass flow rate was assumed due to increased air preheater leakage.

- While no change in ductwork thickness was required, the higher negative pressures required a slight increase in ductwork stiffening steel.

- Incremental cost includes the sum of fan and motor differential costs.

2.3.6 Ammonia Storage, Handling, and Injection System

The following assumptions were used in the development of the ammonia storage, handling, and injection system:

- Anhydrous ammonia delivered by truck.

- One 15,000-gallon tank to provide 10 days of storage at 100 percent boiler load.

- The base case unit utilizes two 100-percent capacity 60-kW electric vaporizers integral to the ammonia storage tank. Higher ammonia consumption due to larger unit size and/or increased NO$_x$ reduction may require the use of steam vaporizers in lieu of electrical vaporizers.

14

- Ammonia will be conveyed to the ammonia injection grid using dilution air. Three 50-percent capacity dilution air fans will provide dilution air at a ratio of 95 percent air / 5 percent ammonia (by volume).

- The ammonia injection grid will include "tunable" zones to allow optimization of the ammonia injection. The number of zones will vary with the required NO_x removal percentage.

2.4 Economic Premises

The base case economic evaluation includes total capital requirement, fixed and variable operating costs, and levelized costs for a new 250-MW pulverized coal utility boiler. Two different sets of economic factors are calculated to permit the economics to be presented either on a current dollar basis, which includes the effect of inflation, or constant dollar basis which ignores inflation. The methodology used to calculate the economic factors is consistent with guidelines established by EPRI in their Technical Assessment Guide (TAG).

2.5 Economic Evaluation Parameters

A detailed list of economic evaluation parameters used to calculate capital charge and levelization factors is presented in exhibit A. The economic parameters assumed for this evaluation, while not company specific, are representative of typical domestic utility financing. Adjustments in economic assumptions were made, since the construction of an SCR in the context of a new plant will have some shared expenses and economies of scale which are not applicable to stand alone retrofit situations. The assumptions include:

- A 30-year plant life was considered applicable for new SCR construction.

- For purposes of calculating the allowance for funds during construction (AFUDC), the SCR construction period was assumed to be 18 months. This resulted in a multiplier of 1.91 percent for the SCR equipment.

- Construction downtime is not applicable for new construction.

- None of the new coal-fired SCR installations in the United States have included a royalty fee to the end user. Therefore, this cost is assumed to be zero.

- All cost data are presented in 1996 dollars. A 3-percent annual inflation rate is assumed for the current dollar analysis.

A DOE spreadsheet model was utilized to compute the capital charge factors and O&M levelization factors. The model, developed by Burns & Roe Services Corporation for utilization by PETC in evaluating a variety of clean coal technologies, utilizes the EPRI TAG methodology in calculating the applicable financial factors. The factors being utilized for this economic evaluation are presented below in table 4.

Table 4
Capital Charge Factors and O&M Cost Levelization Factors

Current Dollar Analysis:

Capital Charge Factor	0.150
O&M Cost Levelization Factor	1.362

Constant Dollar Analysis:

Capital Charge Factor	0.116
O&M Cost Levelization Factor	1.000

2.6 Capital Cost Methodology

The capital cost methodology must reflect all utility costs incurred (including incremental costs) and address a complete scope of supply for a commercial SCR system. For example, the differential cost of an ID fan for a unit without SCR, compared to a unit with SCR, is seldom assessed against the SCR scope of the project. This differential cost, while real to the utility, is more commonly assessed to either a fan or draft system account which does not fully capture the economic impact to balance-of-plant systems due to the SCR. Similarly, differential structural steel cost for the SCR portion of a boiler building is small in the context of the overall boiler building and, therefore, is often included in the boiler building scope and not included as an incremental cost in the SCR scope.

In contrast, the capital cost estimates prepared for this economic evaluation include incremental cost adders applicable to new facilities, which are due to incorporation of SCR into the flue gas train. The following elements are typical of the incremental costs included in the capital cost estimates:

- Incremental boiler house structural steel, siding, and roofing.

- Incremental foundation cost to support the SCR structure.

- Incremental air preheater cost (size, weight, coatings, motor, appurtenances).

- Incremental ID fan (fan and motor) cost due to increased volume and static requirements.

- Incremental ash handling/hoppers due to additional ductwork and/or SCR reactor.

Capital costs were developed based on detailed equipment scope estimates and material take-off quantities. Equipment fabrication and erection estimates were developed using industry standard cost estimating techniques. Much of the cost estimating was accomplished using SCS data based on historical plant design projects. Vendor quotes were obtained for components when little or no data were available or when specific incremental costs were needed. Where possible, validation of the SCR estimates using commercially available literature was used in an effort to

16

reinforce confidence of the estimates. The capital cost also reflects lessons learned from the design, construction, and operation of the SCR test facility.

All capital cost estimates are divided into process areas to provide interested parties sufficient detail to modify costs for project-specific or utility-specific analysis. The major process areas used in the capital cost estimate include:

- Catalyst - Catalyst estimates are applicable for either (or both) honeycomb or plate type catalysts. Current market pricing of $400/ft^3 was the basis of the estimate.

- Reactor Housing, Ductwork, Steel - This area includes all scope associated with the reactor housing; straightening grid (dummy bed); economizer bypass; ductwork; dampers; expansion joints; structural steel; foundations; access platforms; grating; insulation; and flow model study.

- Sootblowers - All catalyst layers include rake-type, retractable steam sootblowers complete with steam piping, valves, insulation, hangers, and steam traps.

- Ammonia Storage, Handling, and Injection - This scope is associated with receiving, storing, handling, and injecting of anhydrous ammonia which includes pressure vessel storage tank(s); steam and/or electric vaporizer; truck unloading facilities; civil works; water deluge/fire protection system; dilution air fans; ammonia piping; valves; multiple-zone, in-duct injection piping; and safety equipment.

- ID Fan Differential - This area reflects the cost differential (fan and motor) due to the increased volume and static pressure duty caused by the SCR. All fans are directly connected with inlet vane control. The SCR will not impact the costs of variable speed drives or fluid couplings.

- Air Preheater Differential - This area captures the incremental cost differences, such as basket material changes, of a deNO$_x$ air preheater relative to a non-SCR application air preheater. Enamel coating of intermediate and cold end heat transfer surface; additional steam sootblowers and water washing equipment; and upgrading of support bearing due to larger, heavier air preheater are other such incremental cost differences which might be present or anticipated.

- Ash Handling Differential - The ammonia slip from a new SCR reactor is assumed to be controlled to prevent unacceptable ammonia-in-flyash contamination and that no flyash treatment systems for ammonia removal will be required. This area captures incremental equipment and pneumatic handling systems present or necessary because of additional ash hoppers either on ductwork or the reactor housing.

- Electrical - The SCR is assumed not to significantly impact the station service transformer or switchyard for a new unit. This area captures the incremental electrical scope (motor

17

control centers, motor starters, cable, conduit, cable tray, etc.) associated with the SCR equipment only.

- Instrument & Controls - This area includes additional instruments such as pressure, flow, and temperature transmitters; wiring; and increased I/O on the plant control system. Inlet and outlet NO_x and O_2 measurement equipment (separate from the plant CEM system) is also included.

- Testing, Training, Commissioning - Costs associated with startup, commissioning, optimization testing, contract acceptance testing, and plant personnel training are included.

Indirect Costs

General facilities are typically calculated as a percentage of total process capital (TPC). A multiplier of 2 percent was assumed for the SCR portion of the general facilities.

Similar to general facilities, engineering and home office fees are also calculated as a percentage of TPC. Because the catalyst is a significant percentage of the capital cost typically engineered by the catalyst supplier, a multiplier of 8 percent of TPC was assumed. This multiplier was an effort to ensure engineering cost is not improperly applied to large, subcontracted items not typically in the scope of the architect/engineer.

The project contingency factor utilized in this evaluation indicates the level of confidence in the total process capital cost of the SCR scope. Project contingencies for a new plant case are lower in comparison to retrofit applications where additional complexity and unknowns related to equipment demolition and relocation can have a significant cost impact.

Project contingency factors are somewhat subjective and reflect an element of uncertainty in both the estimate accuracy and the application of the technology. An overall project contingency of 15 percent was assumed due to offsetting circumstances. High contingency factor circumstances include process uncertainty and operational problems on high-sulfur fuels. Low contingency factor circumstances include unencumbered new unit construction, an estimate that is better than conceptual but not as good as budget, and test facility data.

Preproduction costs as defined by EPRI represent a 1-month total O&M cost (fixed + variable) plus miscellaneous cost items prior to commercial plant operation. For purpose of this evaluation, preproduction costs were estimated using a simpler procedure (as recommended in the DOE guidelines document) in which the total monthly O&M expenses are multiplied by the number of months of anticipated startup operation before the commercial operation date.

$$\text{Preproduction Cost} = \frac{\text{total O\&M costs (less by-product credit) x 2 months of startup}}{12 \text{ months}}$$

18

Inventory capital is calculated as the value of 60 days expendable commodities (assumed to be the variable O&M portion) as defined by the following relationship:

$$\text{Inventory Capital} = \frac{\text{variable O\&M cost (less by-product credit)} \times 60 \text{ days}}{365 \text{ days}}$$

Initial catalysts and chemicals are assumed to be zero, since the initial catalyst volume is captured in the direct capital cost and the ammonia, lubricants, and expendables are captured under preproduction and inventory capital costs.

2.7 Operation and Maintenance (O&M) Costs

Fixed O&M costs include estimates of operating labor, maintenance labor, administration/support labor, and maintenance material. Operating labor costs are calculated as the product of the number of operators per shift, the total operating hours per year, and the operating labor pay rate. It was assumed that the SCR would require one plant equipment operator per shift working half-time. The unit labor man-hour rates are included in the fixed and variable O&M assumptions shown in table 5.

Consistent with EPRI's methodology, total maintenance cost is calculated as a percentage of the total process capital and then apportioned between maintenance labor and maintenance material. For processing liquids and gases, a multiplier of 2 percent was used to determine total maintenance. The proportions of 40 percent and 60 percent, respectively, were used to calculate maintenance labor and maintenance material.

Administrative and support labor is calculated as 30 percent of the total operating and maintenance labor costs.

Variable O&M captures the cost of all commodities as well as costs of expendables such as anhydrous ammonia, catalyst addition/replacement, and utilities. Variable O&M also includes the boiler efficiency penalty incurred due to increased APH outlet gas temperature. Because variable O&M costs are dominated by catalyst replacement, the catalyst management plan is one of the most significant factors affecting overall costs of SCR technology. As noted above, a catalyst guarantee life of 2 years along with deactivation data as measured in the test facility were used to determine the catalyst management plan.

There are two possible sources of heat rate (boiler efficiency) penalty due to the application of SCR to a high-sulfur coal unit. The first, which is included in the O&M costs, is due to the increase of SO_3 in the flue gas which results in a higher acid dew point and corresponding higher air preheater outlet gas temperature. Based on the design criteria of 0.75 percent oxidation of SO_2 to form SO_3 across the catalyst, the resulting increase in air preheater outlet gas temperature is approximately 10°F. The penalty was estimated as incremental fuel burned due to loss in boiler efficiency.

The second source of heat rate penalty, which is not included in this estimate, is due to the required operation of the SCR at lower boiler loads when bypass of the economizer is necessary

to maintain the minimum operating temperature of the SCR. Because it is assumed that the new unit dispatches as a base load unit, it is anticipated that this penalty will be small. However, as evidenced in the retrofit section where the cycling pattern of a given unit is known, the economic penalty for low-load operation can be significant.

Table 5
Fixed and Variable O&M Assumptions and Unit Costs

SCR inlet NO_x	0.35 lb/MBtu
SCR reduction efficiency	60%
Anhydrous ammonia cost	$250/ton
SCR catalyst cost	$400/ft^3
SCR catalyst guarantee period	2 years
SCR catalyst escalation	3.0%
Power cost	30 mills/kWh
ID fan efficiency	75%
SCR draft loss (fully loaded reactor)	3.0 in. W.G.
Ductwork draft loss	0.75 in. W.G.
Ammonia injection grid draft loss	0.75 in. W.G.
Unrecoverable air preheater draft loss	1.0 in. W.G.
Fuel cost (delivered)	$2.00/MBtu
Operating labor man-hour rate	$23.00/hr
Maintenance factor (% of total process capital)	2.0%

The following O&M costs were not included in this evaluation due to the difficulty in estimating the overall impact:

- Catalyst Disposal - Based on current commercial experience, it is assumed that the catalyst supplier would take back spent catalysts.

- Sootblowing Steam - Superheated sootblowing steam required for the SCR was assumed to be small in comparison to the amount used by the boiler.

Table 5 presents the assumptions and unit costs used to calculate the fixed and variable O&M costs for the base case evaluation.

2.8   250-MW Base Case Results

Exhibit D contains detailed results of the capital, O&M, and levelized costs for the 250-MW base case unit. The total capital requirement for a new SCR installation was estimated at $54/kW or $13,415,000 in 1996 dollars. Total first year O&M is $1,045,000 in 1996 dollars. Table 6 summarizes the results shown in exhibit D.

Table 6
250-MW Base Case - New SCR Results

Total Capital Requirement	$ 13,415,000
Total Capital Requirement	$ 54/kW
First Year Fixed Operating Cost	$ 312,000/yr
First Year Variable Operating Cost	$ 733,000/yr
Current Dollar Analysis	
Levelized Cost (mills/kWh)	2.57
Levelized Cost ($/ton NO_x Removed)	$2,500
Constant Dollar Analysis	
Levelized Cost (mills/kWh)	1.85
Levelized Cost ($/ton NO_x Removed)	$1,802

The base case results include some interesting comparisons related to the influence of catalyst cost on the capital and O&M cost of an SCR. The catalyst accounts for approximately 21 percent of the total process capital for the SCR installation. From an O&M perspective, catalyst is approximately 61 percent of the variable O&M and 43 percent of the total annual O&M cost. This result underscores the fact that O&M costs are dominated by catalysts.

When comparing SCR with other NO_x reduction alternatives, the higher capital costs of SCR dominate the levelized cost. For the 250-MW base case, the capital cost is 59 percent of the current dollar total levelized cost, indicating a major portion of the levelized cost is going toward debt service (revenue requirement) of the capital investment rather than operating costs.

21

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3.0   Sensitivity Analyses (Effect of Variables on Economics)

Sensitivity analyses were performed to examine the impact of major process variables on the capital, O&M, and levelized cost of SCR technology. The major sensitivity cases examined as part of this evaluation are summarized in the following sections.

3.1   Capital, O&M, and Levelized Cost for New SCR Vs. Unit Size (60 Percent NO_x Removal)

In order to examine the change in SCR costs vs. unit size, additional capital and O&M estimates were prepared for a 125-MW unit and 700-MW unit. To maintain consistency with the 250-MW base case unit, an SCR removal efficiency of 60 percent NO_x reduction was assumed. Where possible, consistent (or identical) assumptions were made with regard to the 125-MW and 700-MW units.

The combustion calculations for a new 125-MW unit are shown in exhibit F. All assumptions used to prepare the combustion calculation are identical to the 250-MW base case unit. The resulting heat input and coal feed are 1,188 MBtu/hr and 95,000 lb/hr, respectively. A single SCR reactor having similar design criteria as the 250-MW base case (shown in table 3) is assumed for the 125-MW size unit.

The combustion calculations for a new 700-MW unit are shown in exhibit H. All assumptions used to prepare the combustion calculation are identical to the 250-MW base case unit. The resulting heat input and coal feed are 6,650 MBtu/hr and 532,000 lb/hr, respectively. Due to the size of the 700-MW unit, it is assumed that the draft train is split into two 50-percent capacity SCR reactors, each one having similar design criteria as the 250-MW base case (shown in table 3). Two air preheaters are assumed for the 700-MW unit.

When plotted in $/kW vs. unit size, the total capital requirement of the SCR system shows a trend of decreasing unit cost with increasing unit size, indicating significant economy of scale. Total capital requirement ranges from $61/kW for the 125-MW unit to $45/kW for the 700-MW unit. Figure 4 shows a graphical representation of total capital requirement ($/kW) vs. unit size.

Figure 5 shows the levelized cost ($/ton) vs. unit size. The levelized cost decreases with increasing unit size due, in part, to larger NO_x tonnages removed; however, the trend does not appear to be overly sensitive to unit size.

Tabular results showing capital, O&M, and levelized cost vs. unit size for an SCR with 60 percent NO_x removal efficiency are summarized in table 7.

Total Capital Requirement vs Unit Size
SCS/DOE Project – Economic Evaluation
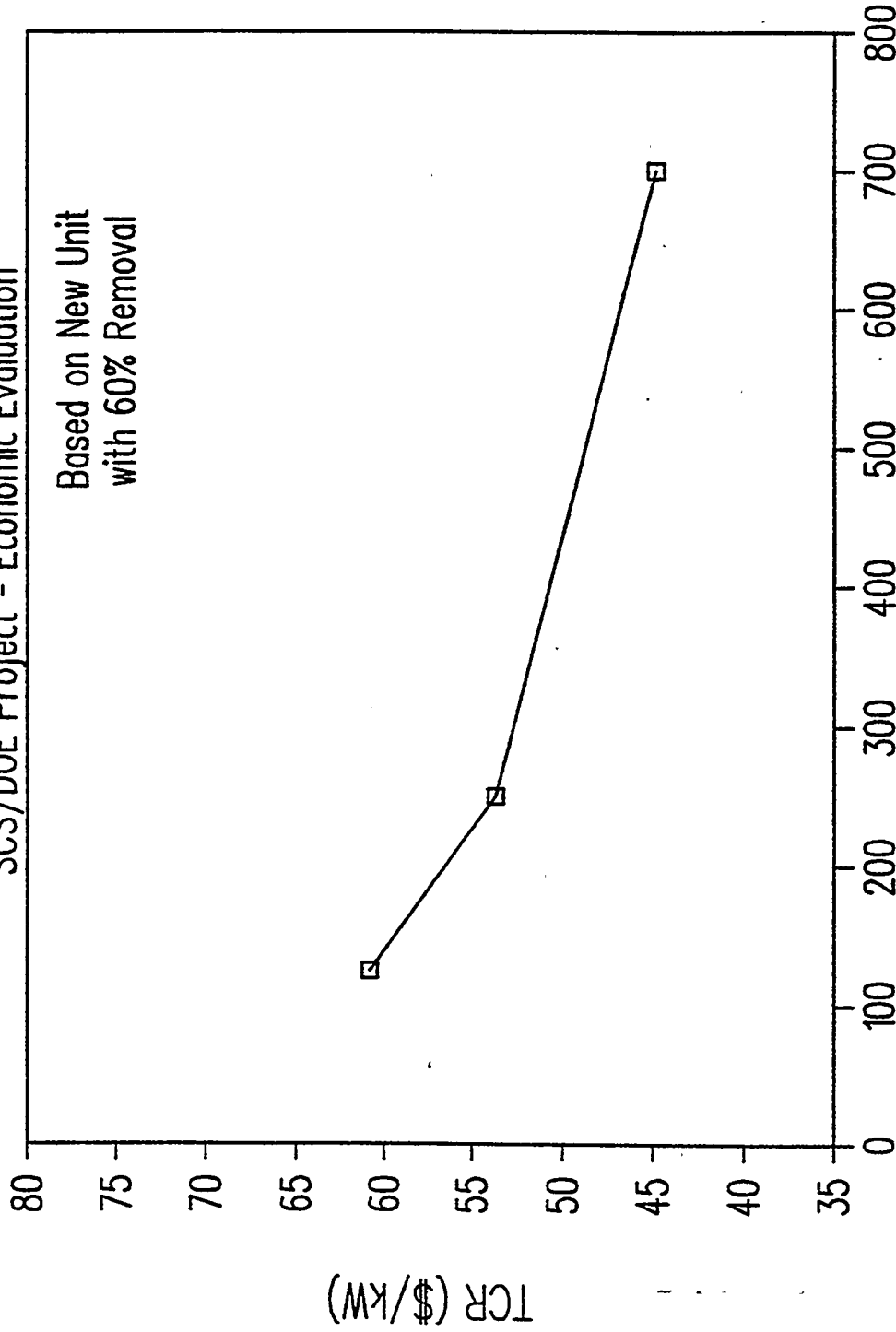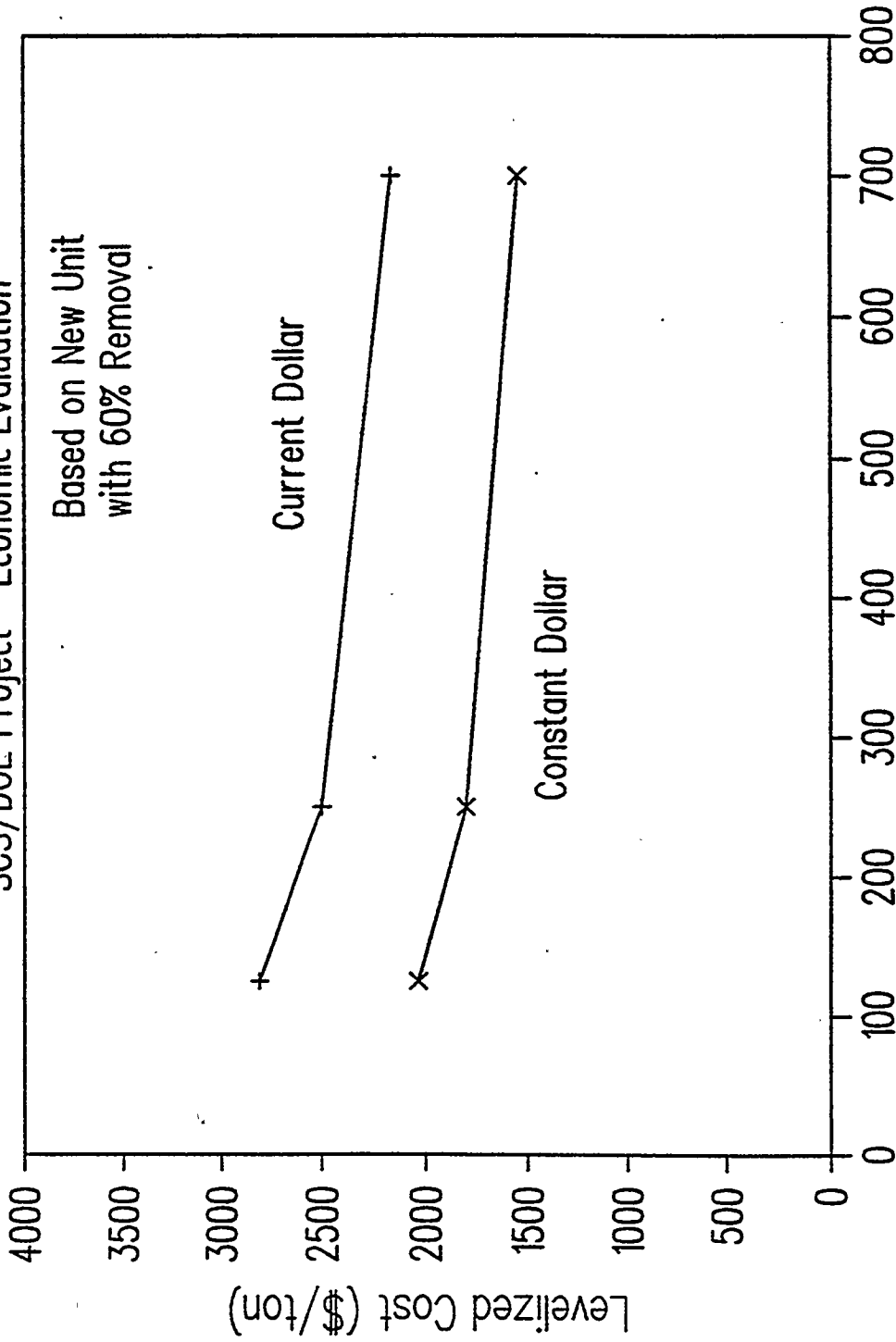
Based on New Unit
with 60% Removal

Unit Size (MW)

Figure 4

TCR ($/kW)

DOESIZE2.atb

24

Levelized Cost vs. Unit Size
SCS/DOE Project – Economic Evaluation

Based on New Unit
with 60% Removal

Current Dollar

Constant Dollar

Unit Size (MW)
Figure 5

Levelized Cost ($/ton)

DOESIZE3.atb

Table 7
Capital, O&M, and Levelized Cost for New SCR Vs. Unit Size
(60 Percent NO$_X$ Removal)

| | Unit Size (MW) | | |
| | | Base Case | |
	125-MW	**250-MW**	**700-MW**
Total Capital Requirement	$7,602,000	$13,415,000	$31,327,000
Total Capital Requirement	$ 61/kW	$54/kW	$45/kW
First Year Fixed Operating Cost	$ 213,000	$312,000	$614,000 --
First Year Variable Operating Cost	$ 367,000	$ 733,000	$2,053,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.89	2.57	2.22
Levelized Cost ($/ton)	$2,811	$2,500	$2,165
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	2.09	1.85	1.59
Levelized Cost ($/ton)	$2,037	$1,802	$1,547

Exhibits G and I include capital, O&M, and levelized cost summaries for the 125-MW and 700-MW units, respectively. The 250-MW base-case unit summary is included in exhibit D.
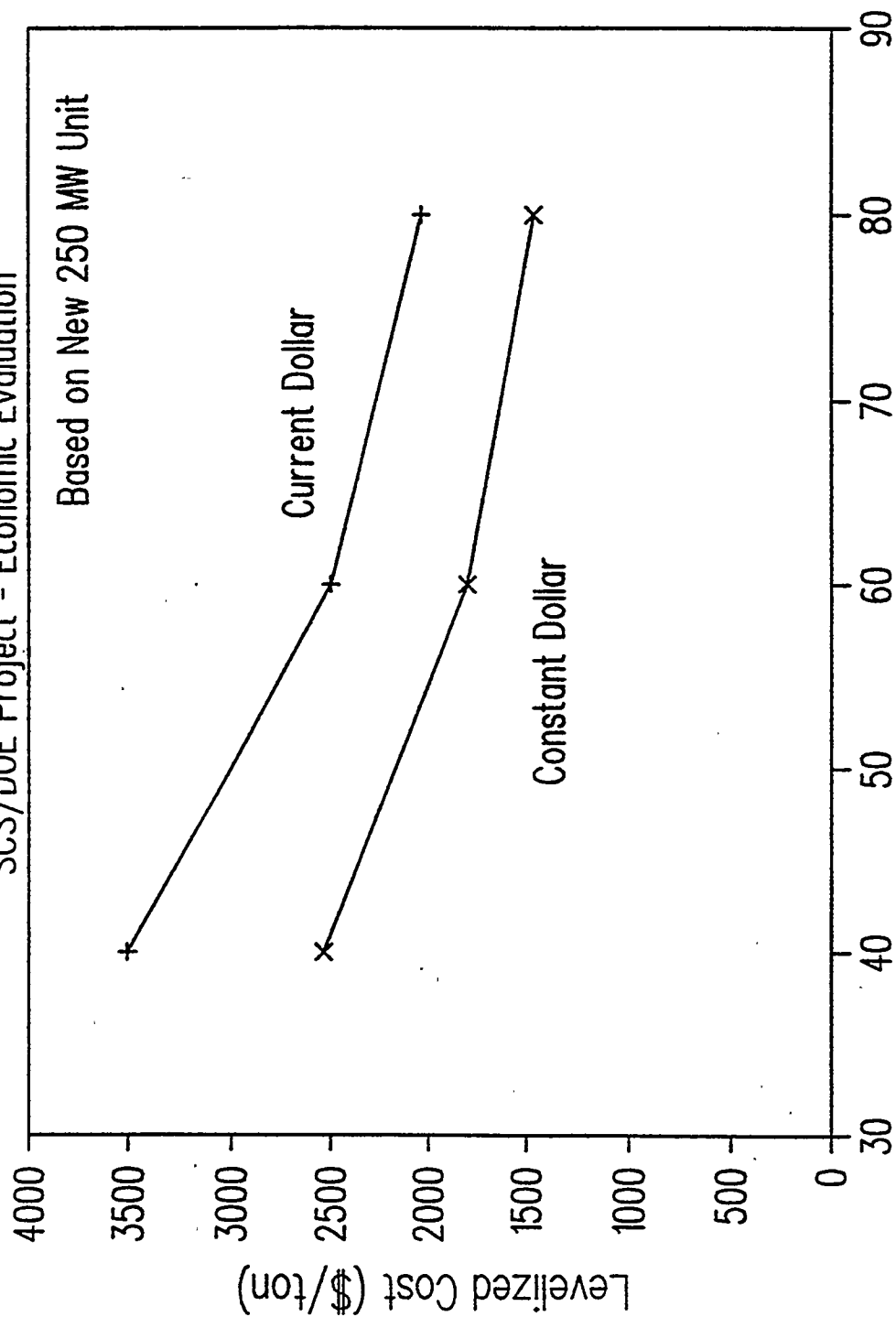
3.2 Capital, O&M, and Levelized Cost for New SCR Vs. NO$_X$ Removal Efficiency (250-MW Plant Size)

In addition to the 250-MW base case with a NO$_X$ removal efficiency of 60 percent, two additional NO$_X$ removal cases at 40 percent and 80 percent were calculated to examine the impact of levelized cost vs. NO$_X$ removal efficiency.

Figure 6 shows the levelized cost ($/ton) vs. NO$_X$ removal efficiency. The levelized cost decreases with increasing NO$_X$ removal percentages and is fairly sensitive to the percentage removal. Thus, once committed to an SCR, significant levelized cost savings ($/ton) can be realized for an incremental increase in capital cost. As seen in table 8, the incremental capital cost difference between 40 percent and 80 percent removal is $1,168,000 or approximately a 9 percent increase in capital cost over the 40 percent design. The corresponding difference in current dollar levelized cost is $1466/ton, a 58 percent decrease in $/ton cost from the 80 percent case compared to the 40 percent case. This difference is primarily due to the increased number of tons removed at 80 percent vs. 40 percent.

This trend in lower levelized cost is also very evident in high-NO$_X$ emitting boilers where similar NO$_X$ removal designs (as a percentage) yield lower $/ton due to a larger number of tons removed.

Levelized Cost vs. NOx Removal
SCS/DOE Project – Economic Evaluation

Based on New 250 MW Unit

Current Dollar

Constant Dollar

NOx Removal (%)
Figure 6

Levelized Cost ($/ton)

DOEREM.atb

Tabular results showing capital, O&M, and levelized cost vs. NO$_X$ removal efficiency for a 250-MW unit are summarized in table 8.

Table 8
Capital, O&M, and Levelized Cost for New SCR Vs. NO$_X$ Removal Efficiency
(250-MW Plant Size)

| | NO$_X$ Removal Efficiency | | |
| | Base Case | | |
	40%	60%	80%
Total Capital Requirement	$12,974,000	$13,415,000	$14,142,000
Total Capital Requirement	$52/kW	$54/kW	$57/kW
First Year Fixed Operating Cost	$305,000	$312,000	$324,000
First Year Variable Operating Cost	$621,000	$733,000	$857,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.39	2.57	2.79
Levelized Cost ($/ton)	$3,502	$2,500	2,036
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	1.74	1.85	2.00
Levelized Cost ($/ton)	$2,536	$1,802	$1,460

Exhibits C, D, and E include 250-MW unit capital, O&M, and levelized cost summaries for the 40 percent, 60 percent, and 80 percent NO$_X$ removal cases, respectively.

3.3 Levelized Cost for New SCR Vs. Inlet NO$_X$ Concentration (250-MW Plant and 60 Percent NO$_X$ Removal)

Many new boiler installations face difficult decisions on how to best optimize overall NO$_X$ reduction requirements using a combination of a low-NO$_X$ combustion system and SCR. While maximizing combustion NO$_X$ reductions can allow lower SCR variable O&M, it may have a negative impact to plant cycle efficiency due to increased LOI in the flyash. Increased carbon monoxide production may also be a limiting factor during deep staged combustion. Optimizing the combustion system to minimize LOI can lead to higher NO$_X$ concentrations entering the SCR and, therefore, higher variable O&M costs to achieve a permitted outlet NO$_X$ emission limit.

The relationship between levelized cost ($/ton) vs. SCR inlet NO$_X$ concentration shown in figure 7 indicates a significant trend of increasing levelized cost with decreasing inlet NO$_X$ concentration. In this case, fewer tons of NO$_X$ are removed by the SCR, highlighting a key difference in cost effectiveness between a controlled new unit application and an uncontrolled (or higher NO$_X$ emitting) retrofit application. A constant 60-percent NO$_X$ removal percentage was applied to all inlet NO$_X$ concentrations shown in figure 7.

28

Levelized Cost vs. SCR Inlet NOx
SCS/DOE Project – Economic Evaluation

Based on New 250 MW Unit
with 60% Removal

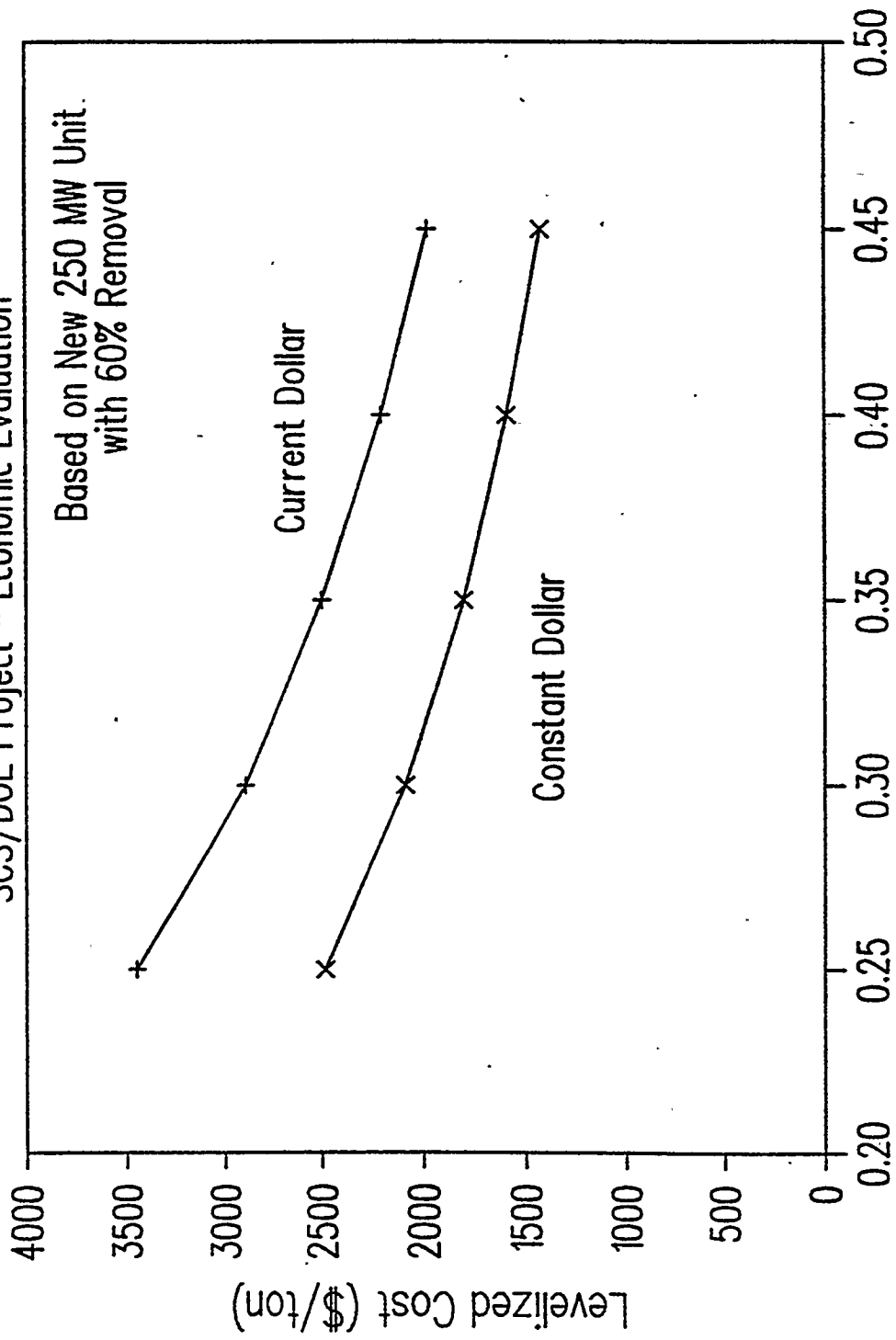Current Dollar

Constant Dollar

Inlet NOx (lb/MBTU)
Figure 7

DOENOIN.atb

Tabular results showing levelized cost vs. SCR inlet NO_X concentration for a 250-MW unit operating at 60 percent NO_X removal are summarized in table 9.

Table 9
Levelized Cost for New SCR Vs. SCR Inlet NO_X Concentration
(250-MW Plant and 60 Percent NO_X Removal)

Inlet NO_X Concentration (lb/MBtu)

Base Case

	0.45	0.40	0.35	0.30	0.25
Current Dollar Analysis					
Levelized Cost (mills/kWh)	2.61	2.59	2.57	2.55	2.53
Levelized Cost ($/ton)	$1,977	$2,205	$2,500	$2,894	$3,446
Constant Dollar Analysis					
Levelized Cost (mills/kWh)	1.88	1.87	1.85	1.84	1.82
Levelized Cost ($/ton)	$1,425	$1,590	$1,802	$2,086	$2,483

Detailed summary sheets for levelized cost as a function of inlet NO_X are included in exhibit J.

3.4 Levelized Cost for New SCR Vs. Catalyst Relative Activity (Catalyst Management Plan) (250-MW Plant and 60 Percent NO_X Removal)

One of the key results produced at the test facility is the catalyst deactivation data collected over the duration of the test program. The variation among selected catalyst deactivation data has been correlated in an attempt to create a range of catalyst management strategies for evaluation.

Relative activity (k/ko) is defined as the activity of the catalyst at a given operating time, k, divided by the activity of the new catalyst at time zero, ko. As noted previously, catalyst deactivation measurements were periodically taken by removing catalyst samples from the reactors and returning the samples to the respective catalyst supplier for analysis. Each catalyst supplier performed a standard protocol of laboratory and bench scale tests to determine the k/ko relationship vs. time for their respective catalyst. Because the relative activity data indicate a wide variation in values as well as the fact that each catalyst supplier extracted an unequal number of catalyst samples at different times intervals over the test period, three sets of individual catalyst data were identified for further evaluation.

All of the catalyst management plans included in this evaluation are based on a 16,000 hour (2-year) catalyst life guarantee period. Because there is very little data beyond 8000 operating hours, development of a relationship showing decline of relative activity over time must be extrapolated to some extent. The catalyst management plans developed for sensitivity evaluation are based on k/ko data having values of 0.90, 0.80, and 0.70 after 16,000 hours.

30

Figures 8, 9; and 10 set forth the declining k/ko relationship over time that results in a value after 16,000 hours of 0.90, 0.80, and 0.70, respectively. Taken alone, each figure appears to have a reasonable relationship based on an exponential decline over time. Given the scarce amount of data beyond 8,000 hours, and the differences in specific catalyst selected, any of the k/ko scenarios in figures 8 through 10 are reasonably plausible and are equally likely to occur. Figure 11 shows k/ko data plotted for all catalysts used in the test program, with the three exponentially declining curves overlaid on the data. The range of curves set the limit for the upper (k/ko = .90) and lower (k/ko = .70) bounds of the relative activity variation.

The base case catalyst management plan was selected with a k/ko value of 0.80 for several reasons:
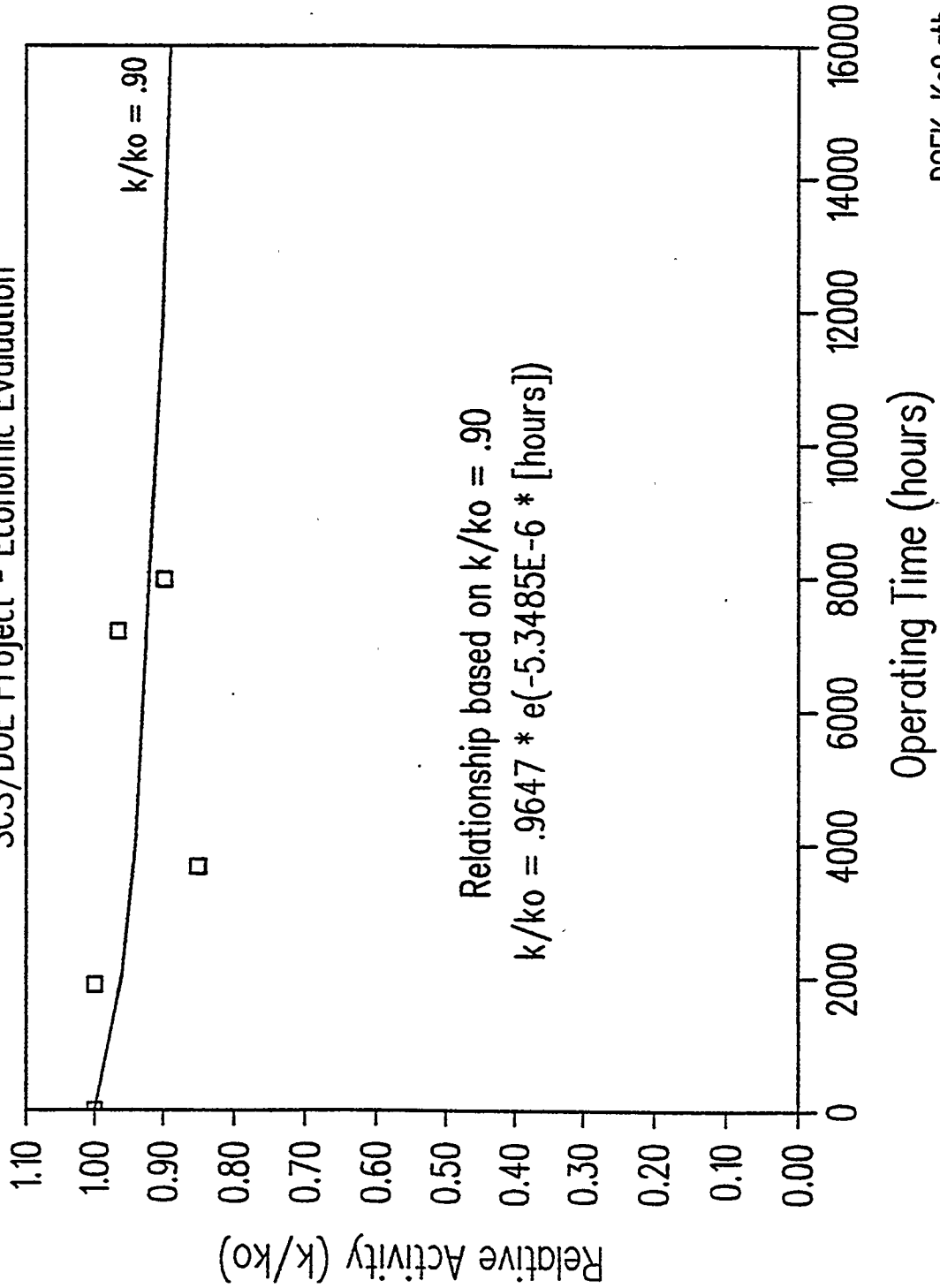
- Figure 9 shows the relative activity relationship for a particular catalyst having the greatest number of data points past 8000 hours. The data appears to reasonably correlate with a k/ko value of 0.80 after 16,000 hours with very little extrapolation.

- Figure 11 shows that a k/ko value of 0.80 after 16,000 is a reasonable median value over the range of possible values.

- Publicly reported commercial experience at two of the new coal-fired SCR installations in the U.S. appears to support the conclusion reached at the test facility that a relative activity of 0.80 after 2 years of operation is reasonable and is representative of U.S. commercial applications as of this writing.

Using the relative activity data, catalyst management plans were developed which define the catalyst replacement schedule over the project life. Knowing the volume of catalyst as well as the time which it is added, project cash flows can be developed. Table 10 indicates the project year in which one layer of catalyst is added and/or replaced.

Table 10
Catalyst Management Plan
Project Years to Add and/or Replace One Layer of Catalyst

k/ko = .70	k/ko = .80	k/ko = .90
2	2	2
5	6	12
7	9	17
9	12	23
12	15	
14	18	
16	21	
19	24	
21	27	
23		
26		
28		

31

Relative Activity Data (k/ko) vs. Time
SCS/DOE Project – Economic Evaluation

k/ko = .90

Relationship based on k/ko = .90
k/ko = .9647 * e(-5.3485E-6 * [hours])

Operating Time (hours)
Figure 8

Relative Activity (k/ko)

DOEK_Ko9.atb

32

Relative Activity Data (k/ko) vs. Time
SCS/DOE Project - Economic Evaluation

k/ko = .80

Relationship based on k/ko = .80

k/ko = .9646 * e(-1.2238E-5 * [hours])
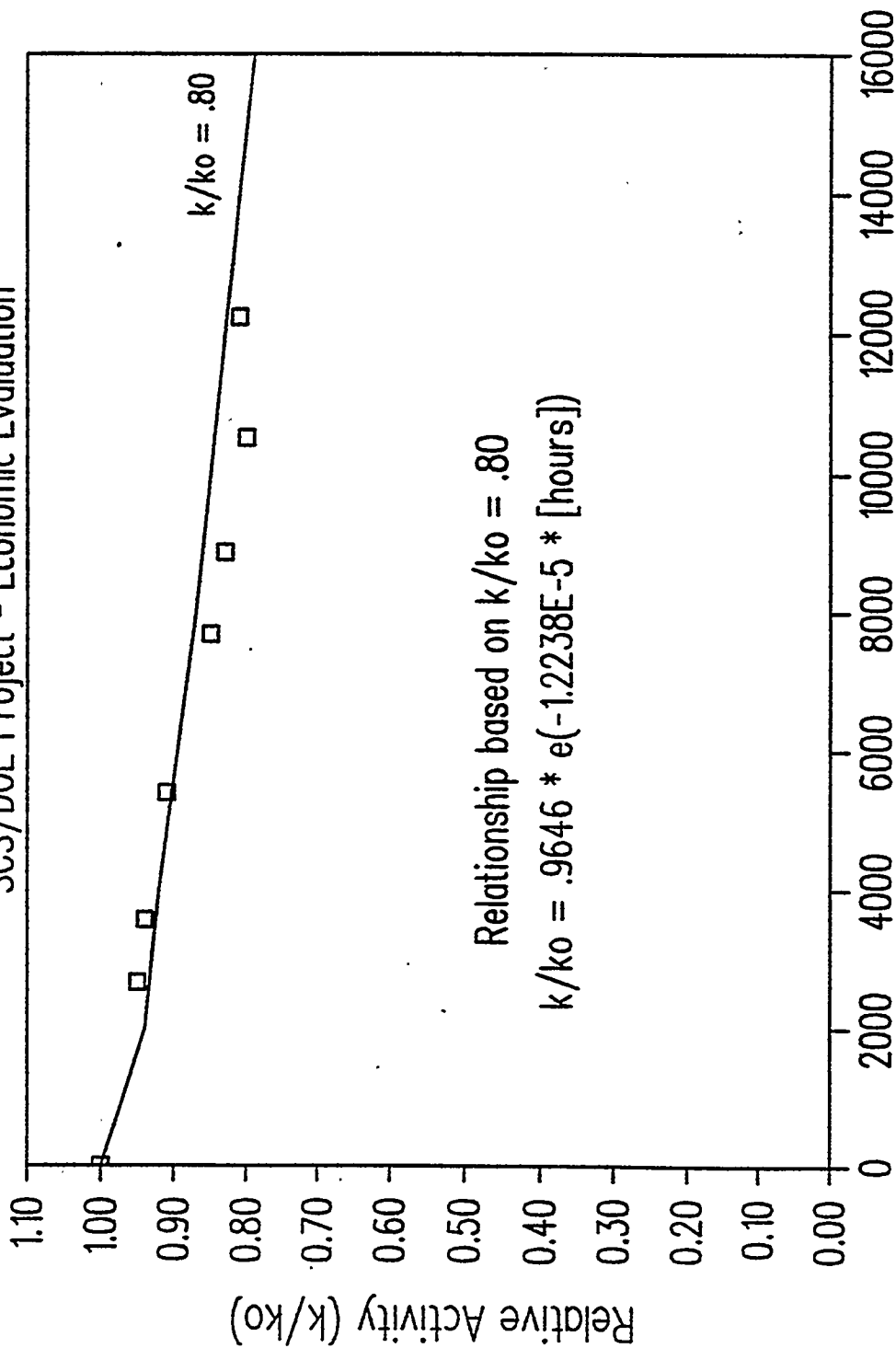
Operating Time (hours)
Figure 9

DOEK_Ko8.atb

Relative Activity (k/ko)

Relative Activity Data (k/ko) vs. Time
SCS/DOE Project - Economic Evaluation

k/ko = .70

Relationship based on k/ko = .70
k/ko = .9734 * e(-2.1501E-5 * [hours])

Relative Activity (k/ko)
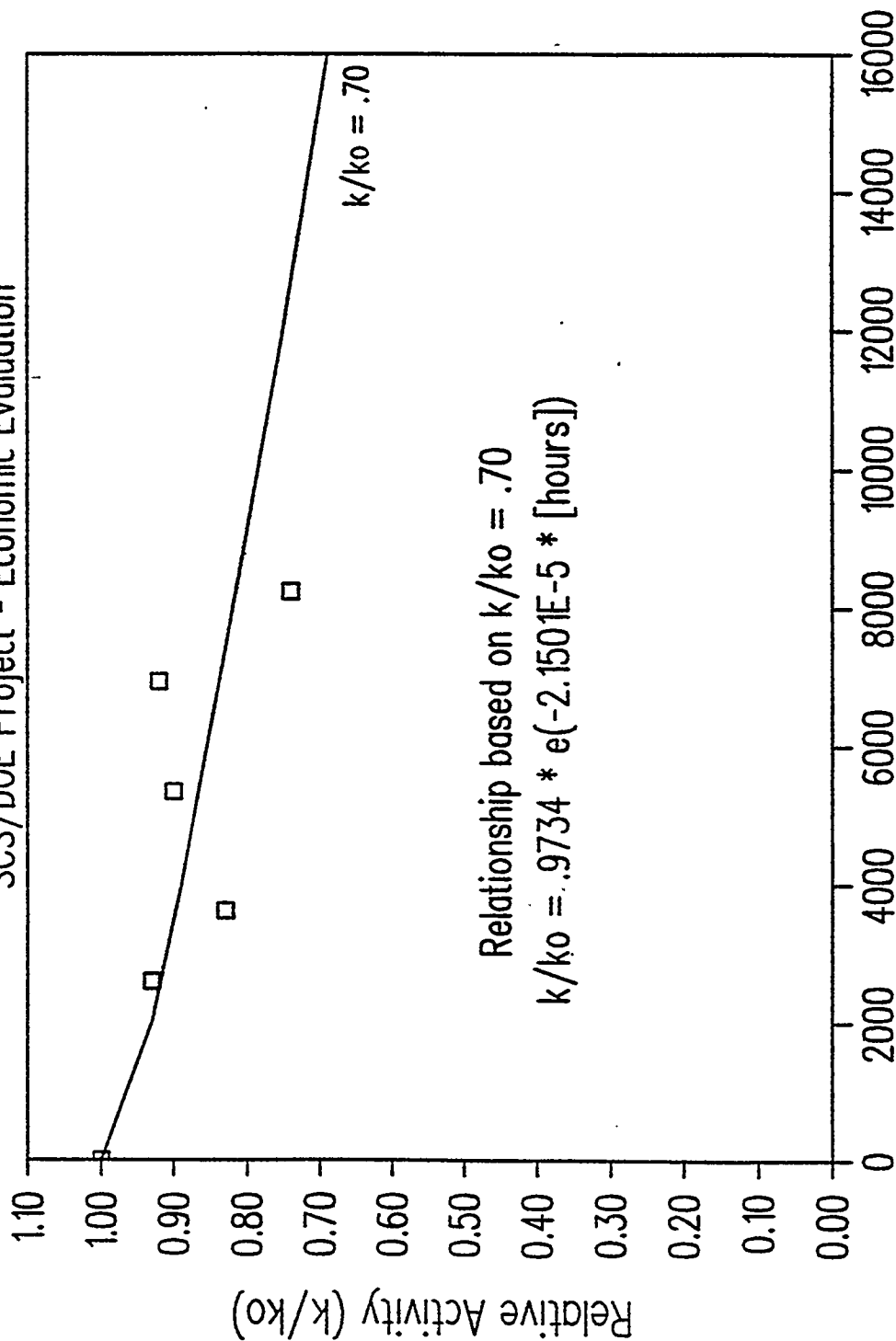
Operating Time (hours)
Figure 10

DOEK_Ko7.atb

34

Relative Activity Data (k/ko) vs. Time
SCS/DOE Project - Economic Evaluation

All Catalyst Deactivation Data

k/ko = .90

k/ko = .80

k/ko = .70

Relative Activity (k/ko)
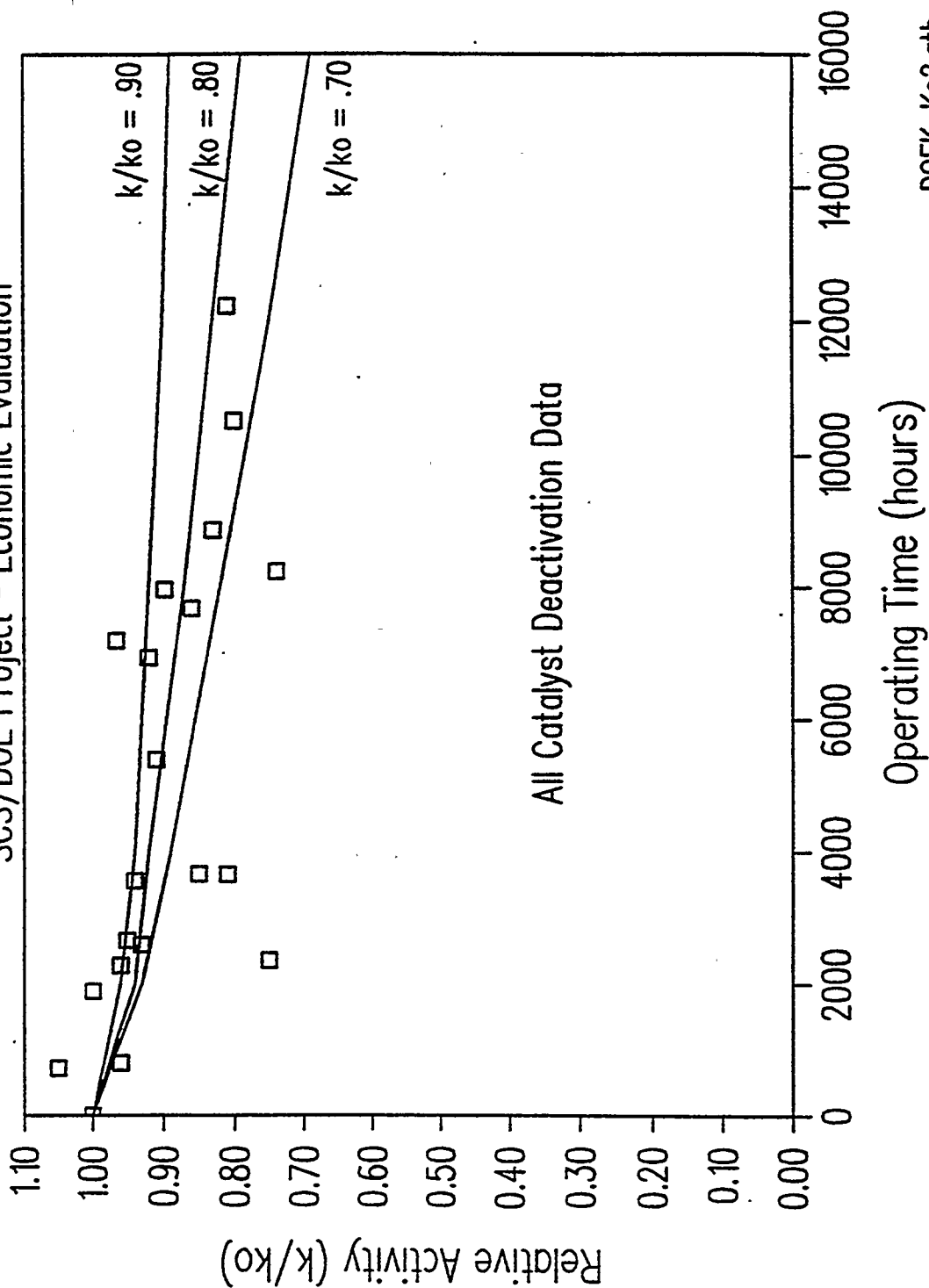
Operating Time (hours)
Figure 11

DOEK_Ko2.atb

35

Table 11 shows the range of total catalyst volume and levelized cost for the three different catalyst management plans. While the total catalyst volume between the most optimistic (k/ko = 0.90) and most pessimistic (k/ko = 0.70) management plans varies by 21,664 ft^3 (a factor of three times), the current dollar levelized cost varies only $373/ton or a difference of only 14 percent (see figure 12). All three catalyst management plans are based on a common 2-year catalyst life guarantee period. So, even though the k/ko = .70 plan adds three times as much catalyst, the catalyst is added in later project years, which has less effect when performing a present value analysis and levelizing to calculate the equivalent annual catalyst cost. This is clearly evident in the fact that the catalyst cost difference between the two cases is $377,000 per year, representing a 64-percent difference in annual O&M dollars (see figure 13). Additionally, if the costs of catalyst disposal were factored in, the expected result would be more pronounced because the k/ko = 0.70 plan would need to dispose of three times as much catalyst.

Detailed summary sheets for levelized cost as a function of different catalyst management plans are included in exhibit K.
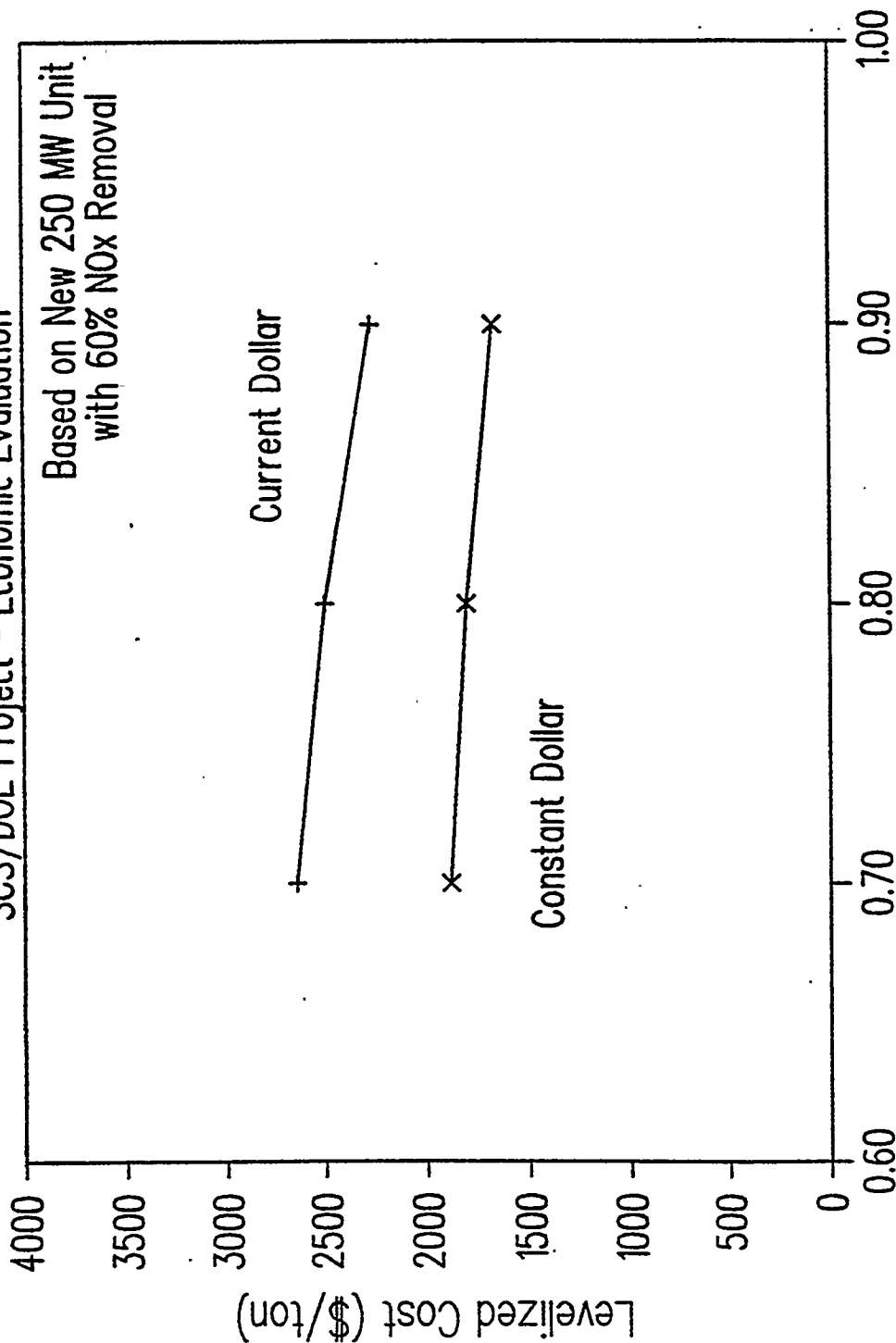
Table 11
Levelized Cost for New SCR Vs. Catalyst Relative Activity
Catalyst Management Plan Results for 250-MW Plant and 60 Percent NO$_X$ Removal

	k/ko = .90	k/ko = .80	k/ko = .70
Total Catalyst Volume Added or Replaced Over the Life of the Plant (ft^3)	10,832	24,372	32,496
Equivalent Annual Current Dollar Catalyst Cost	$216,000	$450,000	$593,000
Equivalent Annual Constant Dollar Catalyst Cost	$144,000	$325,000	$433,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.33	2.57	2.71
Levelized Cost ($/ton)	$2,269	$2,500	$2,642
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	1.72	1.85	1.93
Levelized Cost ($/ton)	$1,671	$1,802	$1,881

It is anticipated that the levelized cost will be very sensitive to timing of catalyst replacement in early years of the project. This is a key issue to consider when contemplating a new project and performing sensitivity analyses to the project pro-forma. Additional efforts might include changing the timing of the catalyst addition as a function of relative activity. For example, the base case management plan based on k/ko = .80 after 2 years would have a positive and negative variant representing a better than expected result (i.e., k/ko = .90 indicates first catalyst addition is actually required after 3 years rather than 2 years) and a worse than expected result (i.e., k/ko = .70 indicates the first catalyst addition is actually required after 1 year rather than 2 years).

Levelized Cost vs. Relative Activity
SCS/DOE Project – Economic Evaluation

Based on New 250 MW Unit
with 60% NOx Removal

Current Dollar

Constant Dollar

Relative Activity (k/ko)
Figure 12

Levelized Cost ($/ton)

4000 3500 3000 2500 2000 1500 1000 500 0

0.60 0.70 0.80 0.90 1.00
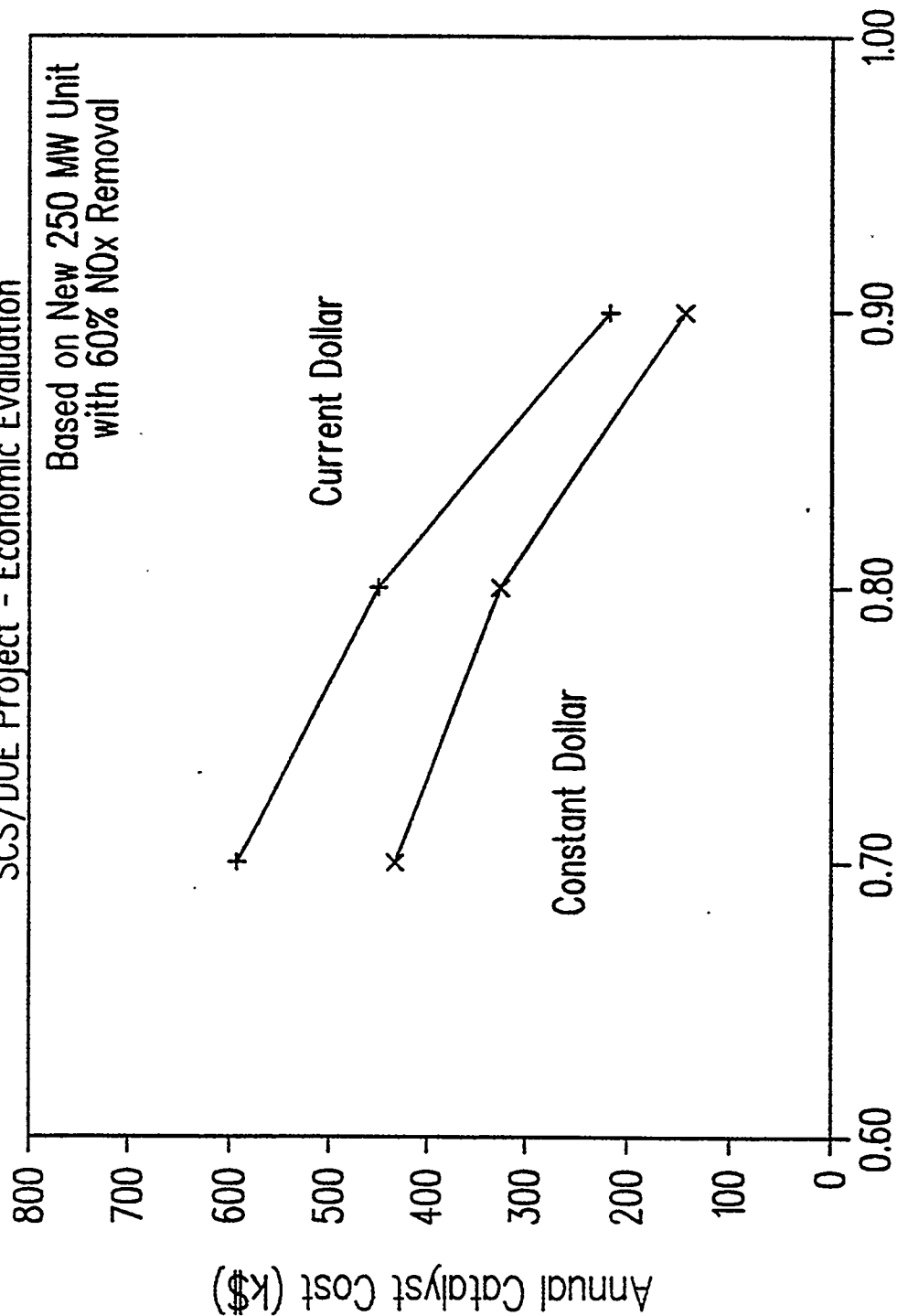
DOECMP.atb

37

Catalyst Cost vs. Relative Activity
SCS/DOE Project – Economic Evaluation

Based on New 250 MW Unit
with 60% NOx Removal

Current Dollar

Constant Dollar

Relative Activity (k/ko)
Figure 13

Annual Catalyst Cost (k$)

DOECMP2.atb

38

3.5 Levelized Cost for New SCR Vs. Return on Common Equity (ROE) (250-MW Plant and 60 Percent NO$_X$ Removal)

The domestic electric utility industry is in transition from a predominantly regulated environment to a more market driven, less regulated environment. There is much uncertainty in the future earning potential of major capital investments such as SCR technology. This regulatory uncertainty is best illustrated by a recent decision in New Hampshire where the Public Utilities Commission disallowed full cost recovery of recently installed SCR technology, thereby impacting the utility's ROE.

Table 12 below summarizes the economic factors and levelized cost as a function of ROE. Figure 14 shows the change in levelized cost as a function of ROE for a 250-MW unit with 60 percent NO$_X$ removal.

As shown in table 12, for every 2-percent increase in ROE, the current dollar levelized cost increases approximately 4.5 percent. Exhibit L contains the detailed calculations of economic factors and levelized cost summaries for a range of ROE values for the 250-MW plant.

Table 12
Levelized Cost for New SCR Vs. Return on Common Equity (ROE)
(250-MW Plant and 60 Percent NO$_X$ Removal)

Return on Equity (ROE)

	Base Case				
	7%	**9%**	**11%**	**13%**	**15%**
Current Dollar Analysis					
Capital Charge Factor	0.132	0.141	0.150	0.160	0.169
O&M Levelization Factor	1.395	1.378	1.362	1.347	1.333
Constant Dollar Analysis					
Capital Charge Factor	0.100	0.108	0.116	0.124	0.133
O&M Levelization Factor	1.000	1.000	1.000	1.000	1.000
Current Dollar Analysis					
Levelized Cost (mills/kWh)	2.36	2.46	2.57	2.69	2.79
Levelized Cost ($/ton)	$2,295	$2,398	$2,500	$2,615	$2,720
Constant Dollar Analysis					
Levelized Cost (mills/kWh)	1.69	1.77	1.85	1.93	2.02
Levelized Cost ($/ton)	$1,646	$1,724	$1,802	$1,880	$1,968

Levelized Cost vs. Return on Equity
SCS/DOE Project - Economic Evaluation

Based on New 250 MW Unit
with 60% NOx Removal

Current Dollar

Constant Dollar
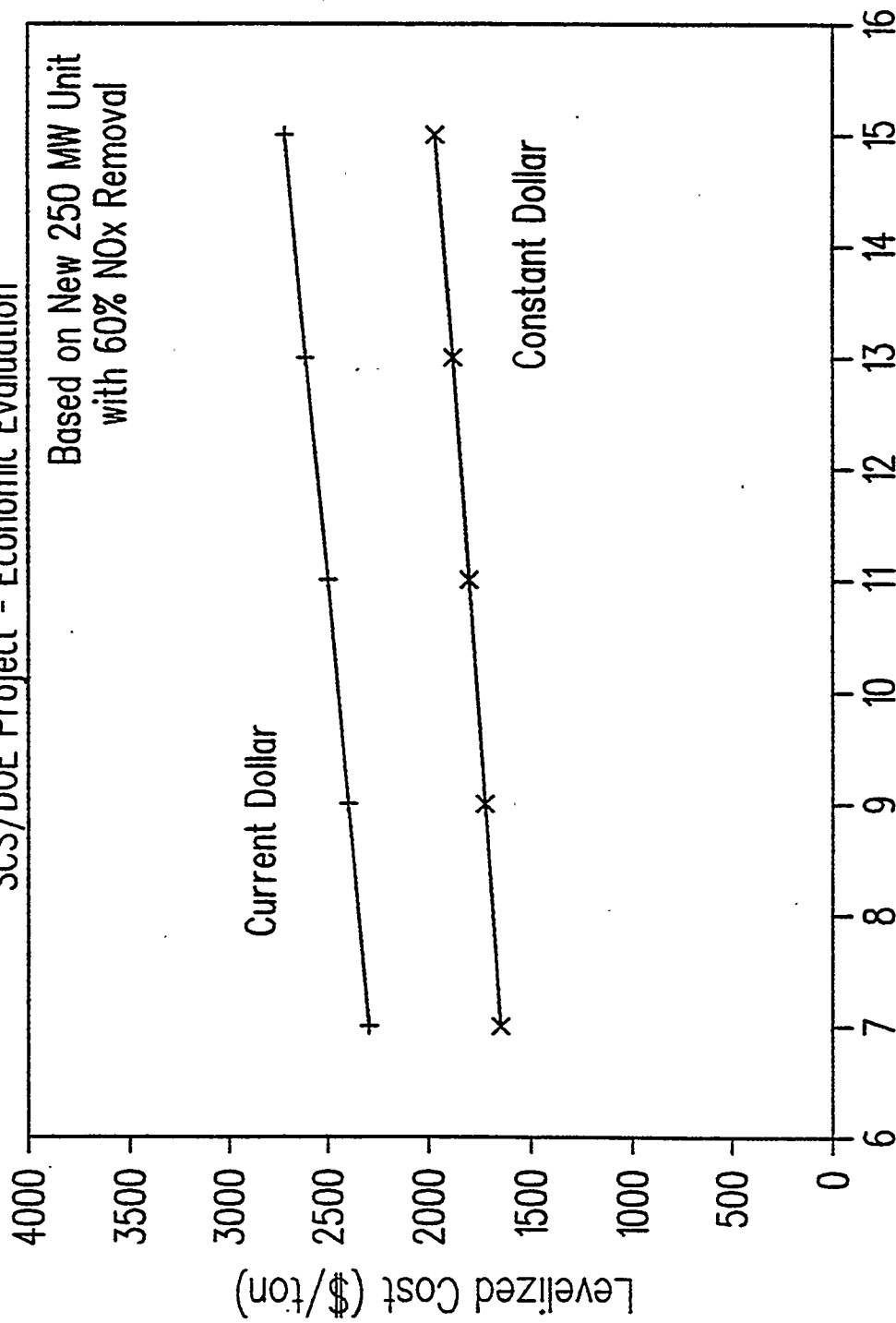
Return on Equity (%)
Figure 14

Levelized Cost ($/ton)

DOEROE.atb

40

3.6 Capital, O&M, and Levelized Cost for New SCR Vs. Catalyst Price (250-MW Plant Size and 60 Percent Removal)

Market price of catalyst can affect both the capital and O&M cost of SCR technology. The most recent experience in Germany during the 1980's resulted in catalyst market price variations ranging between $900/ft^3 and $300/ft^3 over an 8 to 10 year period. More recently in the U.S., one of the five new plants equipped with SCR realized a catalyst price of approximately $400/ft^3.

To address the sensitivity of capital, O&M, and levelized cost to changes in the market price of catalyst, the catalyst price was varied by +/- $50/ft^3. It is recognized that dynamic market forces may cause wider variation in prices than those assumed for this analysis. However, based on the comments of one of the participants of the project, $350/ft^3 was quoted as a realistic, obtainable catalyst price based on current market conditions.

Table 13 shows the capital, O&M, and levelized cost vs. catalyst price for a new 250-MW unit. Varying the catalyst price +/- 12.5 percent·(+/- $50/ft^3) results in a change in levelized cost of only +/- 4 percent. Variable O&M is the most sensitive to changes in catalyst price since catalyst cost dominates this annual expense. Capital cost changed approximately 2 percent since the catalyst represents only 20 to 25 percent of the total process capital.

Table 13
Capital, O&M, and Levelized Cost for New SCR Vs. Catalyst Price
(250-MW Plant and 60 Percent NO$_x$ Removal)

	Catalyst Price ($/ft^3)		
		Base Case	
	$350	**$400**	**$450**
Total Capital Requirement	$13,040,000	$13,415,000	$13,777,000
Total Capital Requirement	$52/kW	$54/kW	$55/kW
First Year Fixed Operating Cost	$306,000	$312,000	$319,000
First Year Variable Operating Cost	$677,000	$733,000	$789,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.46	2.57	2.67
Levelized Cost ($/ton)	$2,398	$2,500	$2,602
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	1.78	1.85	1.92
Levelized Cost ($/ton)	$1,737	$1,802	$1,867

Exhibit M includes capital, O&M, and levelized cost summaries for $350/ft^3 and $450/ft^3 catalyst prices.

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4.0 Application of SCR Technology For a Retrofit Unit

The economic evaluations reported in prior sections of this report were focused on SCR installed on a new coal-fired facility. However, the majority of the near-term future U.S. SCR market may be in retrofit applications. The cost of implementing SCR technology is a topic of considerable debate in the present deliberations by the Ozone Transport Assessment Group (OTAG)[*] and in defining the U. S. Environmental Protection Agency's (EPA) proposed CAAA Title IV NO_x emission limits for Group 2 boilers, specifically cyclones. The boiler types, number of units per boiler type, and existing generating capacity for existing, base loaded duty, coal-fired units in the OTAG region alone are shown in table 14.

Table 14
Boiler Type, Number, and Generating Capacity in the OTAG[*] Region

Boiler Type	Number of units	Generating Capacity, MW
Wall-fired	315	94,327
Tangentially fired	315	112,000
Cyclone	77	22,329
Cell-fired	33	24,143
Wet-bottom	23	4,712
Roof-fired	29	3,111

---

* OTAG has proposed a 37-state nonattainment area encompasing the eastern part of the U.S.

Because of the considerable uncertainty and debate involving SCR retrofit cost for existing plants, SCS has completed a study to determine the cost and technical-feasibility of retrofitting SCR-technology to selected coal-fired generating units within the Southern electric system. While retrofit issues will vary from plant to plant and company to company, the results of this study reflect the typically wide range of retrofit costs due to site-specific issues encountered at those plants studied within the Southern electric system. It is recognized that the costs summarized in this study applicable to The Southern Company may or may not be indicative of other utility's installations due to boiler types, site constraints, and regulatory requirements.

4.1 General Retrofit Issues

From a technical perspective, based on the large number of worldwide applications of SCR on coal-fired boilers, SCR can be judged to be broadly applicable to a variety of boiler types and fuels. It is evident that the engineering issues associated with the design and retrofit of full-scale commercial SCR facilities have been and are being successfully addressed. SCR is more costly, however, when compared to combustion modifications, and exhibits poor economies of scale at smaller boiler sizes. Therefore, applicability and feasibility assessments must also consider site specific economic factors.

Table 15 provides a summary of factors affecting applicability and technical feasibility of SCR when applied to coal-fired retrofit applications.

Table 15
Summary of Factors Affecting Applicability and Technical Feasibility of SCR

FACTOR	COMMENTS
• Coal type and characteristics	• SCR primarily applied commercially to low-sulfur coal. Japanese experience is with clean, washed bituminous coal. European experience is with low-sulfur brown and black coal. Problem with alkali metal poisoning can occur with low-rank coals. Trace metal constituents in coal have potential to be catalyst poisons. Little operational experience on high-sulfur U.S. coals.
• Boiler size	• No limitations.
• Boiler age	• No technical limitations. As with any retrofit technology, remaining useful life affects economics.
• Boiler heat release limitations	• Not applicable to SCR.
• Capacity factor (CF) limitations	• No technical limitations. Prolonged low load helps space velocity of SCR as long as temperature maintained. Low CF hurts economics.
• Load profile	• Uncertain area for SCR. Japanese experience is with baseloaded units. European units will have more cycling duty. More data are needed in this area to assess site-specific impacts.
• Boiler firing type (PC v. cyclone, etc.)	• No technical limitations except that firing type affects flue gas NO_x level. High NO_x levels (>600 ppm) increases the capital cost of SCR. NO_x should first be reduced though combustion modifications, if possible.
• Boiler firing configuration	• Same as above. Tangentially fired boilers have slightly better homogeneous flue gas mixture and lower baseline NO_x than do wall-fired boilers.
• Boiler bottom type (wet v. dry)	• European experience has shown rapid catalyst deactivation on wet-bottom boilers due to fly ash metal vaporization and condensation on catalyst. Wet bottom applications will deactivate faster than dry bottom applications but new arsenic-resistant catalyst improves catalyst life. High-dust SCR widely applied to dry-bottom boilers.
• Geographic applicability	• No limitations, except as might affect shipping costs for NH_3 since major U.S. NH_3 source is U.S. Gulf Coast.
• SCR retrofit difficulty	• Because of the temperature range in which the SCR operates, retrofit feasibility is dictated by having adequate space available to locate large, heavy reactor between the economizer outlet and air preheater inlet.
• Boiler outlet flue gas temperature	• Flue gas temperature variation versus boiler load will dictate extent of economizer bypass required if operation over entire boiler load range is required. Requirement to bypass economizer at low loads will affect unit heat rate and may change heat absorption patterns in the boiler.
• Particulate collector requirements	• Affects type of SCR. "Cold-side" particulate collection (ESP or baghouse) requires high-dust SCR. "Hot-side" ESP allows choice of high- or low-dust SCR. Worldwide, high-dust SCR is the preferred approach.
• Air preheater requirements	• Air preheater physical features influence fouling potential due to ammonium salt formation. Possible degrade in performance due to higher pressure drop and flue gas mass flow. Review of heat transfer surface configuration, material, geometry, orientation, cleanability, temperature profile, leakage, and physical condition should be considered when assessing SCR impacts to existing plant.
• Raw material requirements	• Ammonia and catalyst. No other requirements.
• By-product market limitations	• No salable by-products. Ammonia slip could affect fly ash sales and increase landfill development costs.
• Thermal efficiency penalty	• Thermal penalty possible due to increased air preheater outlet gas temperature with high-sulfur applications. • If SCR is required to operate over wide boiler load range, thermal penalty will be incurred through lower boiler load range (as high as 1.0 percent). Extent of penalty is function of load dispatch.
• Waste disposal factors	• In most cases, spent catalyst is shipped back to catalyst vendor.
• Other factors	• High-dust, hot-side SCR must be considered in its effects on particulate collection efficiency and (if present) effects of slip NH_3 on a downstream FGD process. • Flue gas draft loss across SCR may dictate need for ID fan upgrade or balance draft conversion of boiler.

4.2 Specific Unit Technological Feasibility

The following plant descriptions are the result of inspections made for purposes of formulating a conceptual SCR retrofit design at selected plants within the Southern electric system. SCR performance requirements were estimated using combustion calculations based on field-collected low-NO$_x$ burner acceptance test (or baseline test) information. Conceptual layouts were developed taking into account the retrofit difficulties at each site and the results of the catalyst suppliers'. volume estimates. A material scope was then developed itemizing the major pieces of equipment. Where required, vendor quotes were obtained for required components. Much of the cost estimate was produced using SCS information. All of the units considered under this study are tangentially fired, pulverized coal boilers originally manufactured by Combustion Engineering.

4.2.1 Plant A

Plant A includes two tangentially fired, supercritical units nominally rated at 700 MW each. Each unit has a center wall dividing the furnace into two halves. There are six elevations of coal nozzles in each of the eight corners. The boiler is fired under balanced pressure. Both units are designed to deliver superheated steam at a rate of 5,044,992 lb/hr at turbine inlet conditions of 1000°F and 3500 psig. Both units have been retrofitted with low-NO$_x$ firing systems featuring flame attachment coal nozzles, offset secondary air, and separated over fire air (SOFA).

The SCR arrangement reflects the difficulty in locating the reactors, due to a previously retrofitted cross-over duct for the "piggy back" cold-side precipitators added when the boilers were converted to balance draft operation. The SCR reactors are located directly to the rear of the boiler house, above the existing precipitator ductwork, in an attempt to avoid extensive modifications of the existing precipitator ductwork. This arrangement, as shown in figure 15, produces a "sidewinder" configuration (places the reactors toward opposite sides of the boiler house). This arrangement requires 90-degree horizontal turns in the SCR inlet and outlet duct, but positions the reactor closer to the boiler building with less overhang above the precipitators. Access to the back side of the boiler house wall is relatively unobstructed, but would require structural modifications to the boiler building crossbracing to allow for the ductwork. In addition, possible interference with the coal conveyors, located between Plant A and adjacent units, needs further investigation.

The sidewinder arrangement requires separate support structures for each reactor, but does not require penetration of the precipitator inlet plenum ductwork. A space truss would be used to individually support each SCR reactor, with the column spacing restricted by the available space between the boiler house and the precipitator. In addition, it may be necessary to modify or relocate major foundation, such as ash trenches, U-drains, and small equipment to accommodate this arrangement.

4.2.2 Plant B

Plant B includes two tangentially fired, supercritical units nominally rated at 880 MW each. Each unit has a center wall dividing the furnace into two halves. There are seven elevations of coal
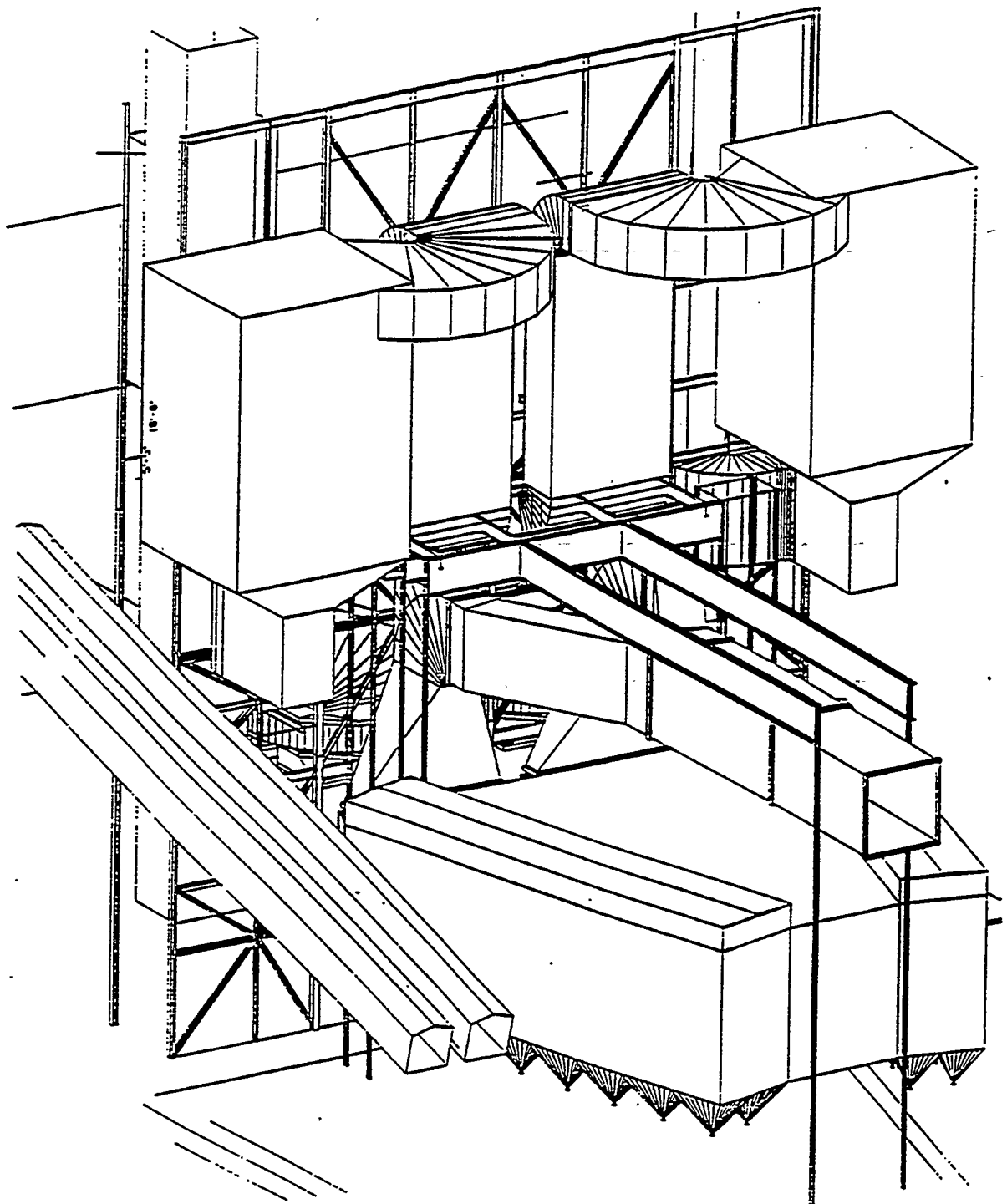
Figure 15
Perspective View of Retrofit SCR Arrangement for Plant A

nozzles in each of the eight corners. The boiler is fired under positive pressure. Both units are designed to deliver superheated steam at a rate of 6,351,470 lb/hr at turbine inlet conditions of 1000°F and 3500 psig. Both units have been retrofitted with low-NO_x firing systems featuring flame attachment coal nozzles, offset secondary air, and SOFA. There are two SOFA boxes, with two compartments each, on each of the eight corners. Each of the two compartments contains tilting air nozzles and individual damper control.

Because Plant B is limited in forced draft (FD) fan capacity during certain periods of the year, it is likely that a draft system upgrade will be required. No attempt has been made to determine if higher positive-pressure operation rather than balance draft conversion is technically feasible. In the absence of a detailed draft study, it is unclear whether or not the pressurized units would require balance draft conversion in order to retrofit SCR. The recent retrofit of an SCR at Public Service of New Hampshire's Merrimack Station illustrates that it is technically possible to retrofit an SCR on a pressurized unit without converting operation to balanced draft. However, this appears unlikely at Plant B due to the already limited capacity of the draft system. In order to bound the financial exposure, the cost estimate includes the balance draft option, which approximately doubles the cost to install SCR.

The proposed arrangement, as shown in figure 16, locates the SCR reactors directly to the rear of the boiler house, above the existing precipitator inlet ductwork. The unit will utilize two SCR reactors. The straight-back configuration of the reactors eliminates the need for horizontal turns in the SCR inlet and outlet ductwork. In figure 17, which shows the side elevation of this arrangement, the economizer outlet duct must turn upward upon exiting the building, thereby causing the reactors to be positioned further outward from the boiler building, above the precipitator inlet plenum duct. Once above the precipitator inlet ductwork, access to the back side of the boiler house wall is relatively unobstructed, but structural modifications will be required to the building crossbracing to allow for the SCR ductwork.

The two SCR reactors parallel the north/south centerline of the boiler building, an arrangement that allows for a common support structure for the two reactors. The proposed support structure for this arrangement consists of four towers supporting a frame common to both reactors. On the south end (toward the boiler building), the frame is supported by two towers and completely spans the precipitator inlet plenum duct. However, on the north end (toward the stack), it is necessary for two support towers to penetrate the plenum. Access to the SCR reactors would be provided on the south side of the reactors and tied into the boiler building.

An alternate plan would be to locate the SCR reactors on the roof of the boiler house. However, there is an existing monorail that extends out approximately 20 ft from the power house wall and runs parallel to the length of the boiler house. The ductwork to and from the SCR would have to be routed beyond the monorail. Because of this, it would not be practical to locate the SCR on the roof. The arrangement described above is located below the existing monorail, which could be used for catalyst additions.
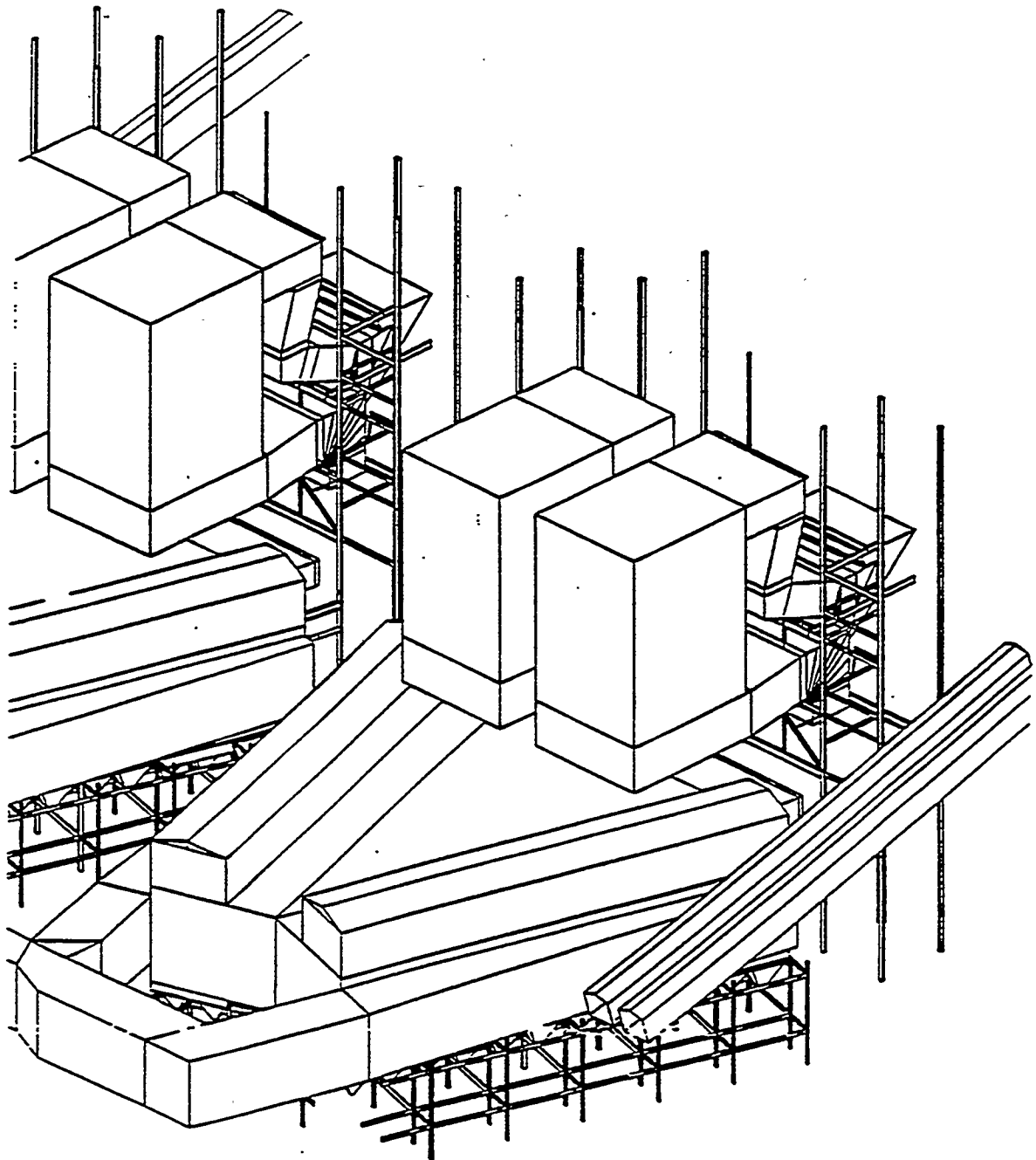
Figure 16
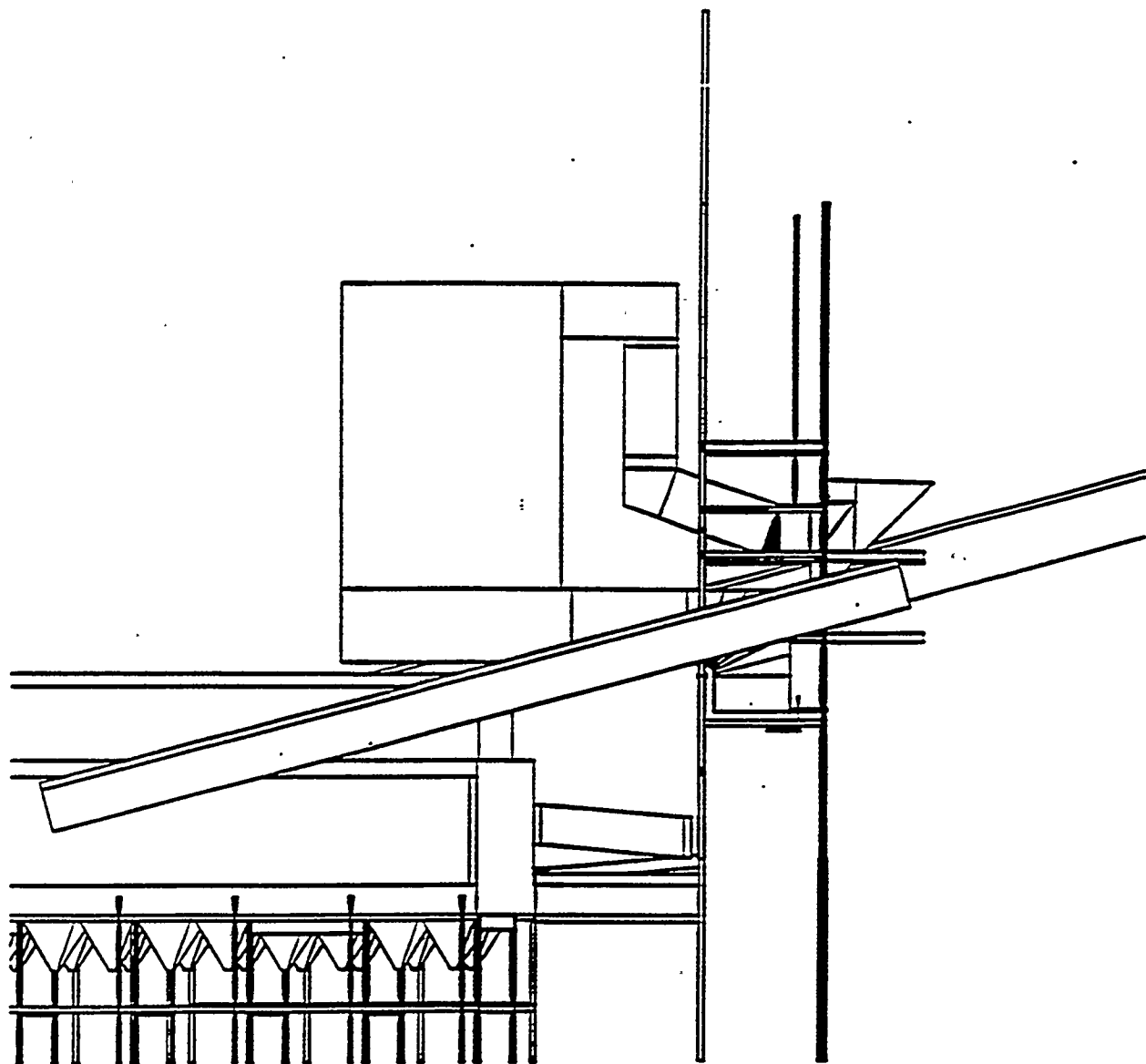Perspective View of Retrofit SCR Arrangement for Plant B

48

Figure 17
Side Elevation View of Retrofit SCR Arrangement for Plant B

4.2.3 Plant C

Plant C includes two tangentially fired, supercritical units nominally rated at 865 MW each. Each unit has a center wall dividing the furnace into two halves. There are seven elevations of coal nozzles in each of the eight corners. The boiler is fired under balanced pressure. Both units are designed to deliver superheated steam at a rate of 6,269,267 lb/hr at turbine inlet conditions of 1000°F and 3500 psig. One unit has been retrofitted with a low-NO_x firing system featuring low-NO_x coal nozzles, offset secondary air, and SOFA. There are two SOFA boxes, on each of the eight corners, with three compartments each of tilting air nozzles.

The proposed arrangement of the SCR reactors is similar to the location shown for Plant B (i.e., above the precipitator inlet plenum duct). Each unit is equipped with chevron-type electrostatic precipitators which have a low profile that provides a clear and unobstructed space above the inlet ductwork. The economizer outlet duct would exit the boiler room wall above elevation 855 ft. The SCR outlet duct would enter the building above elevation 834 ft. A space of approximately 20 ft between boiler house column line and the SCR reactor is needed to allow for moving replacement economizer sections up and into the boiler room.

The two SCR reactors parallel the north/south centerline of the boiler building, allowing an arrangement with a common support structure for the two reactors. The assumed support structure for this arrangement is similar to Plant B in that it will be necessary to penetrate the precipitator inlet plenum duct to support the SCR reactors. Access to the SCR reactors would be provided on the boiler house side, utilizing the existing monorail. Because the forced-draft fan intakes are located on the boiler house roof, the roof is not a viable location for the SCR reactors.

4.2.4 Plant D

Plant D includes two tangentially fired, subcritical units nominally rated at 245 MW each. Each unit has a center wall dividing the furnace into two halves. There are five elevations of coal nozzles in each of the eight corners. The boiler is fired under balanced pressure. Both units are designed to deliver superheated steam at a rate of 1,734,000 lb/hr at turbine inlet conditions of 1000°F and 2400 psig. Both units have been retrofitted with a low-NO_x firing system which features a split-flame, wall-fired low-NO_x burner technology into a corner-fired tilting burner technology. A close coupled overfire air compartment is located above the top coal elevation.

Based on the results of the study, retrofit of SCR would be difficult at these units due to the existing location of previously retrofitted precipitators. The precipitators are elevated over an active railroad spur, which forced the entire precipitator assembly to be displaced vertically upward. The precipitator outlet ductwork is routed over the top of the precipitator, up the back side of the boiler building, and up to the roof. As a result, the precipitators and ductwork effectively block the back side of the boiler house where ductwork tie-ins between the economizer outlet and the air preheater inlet would be required in order to add SCR.

The proposed location of the SCR reactors is on either side of the existing boiler house, as shown in figures 18 and 19. This arrangement (opposite hand for each unit) produces a "sidewinder"
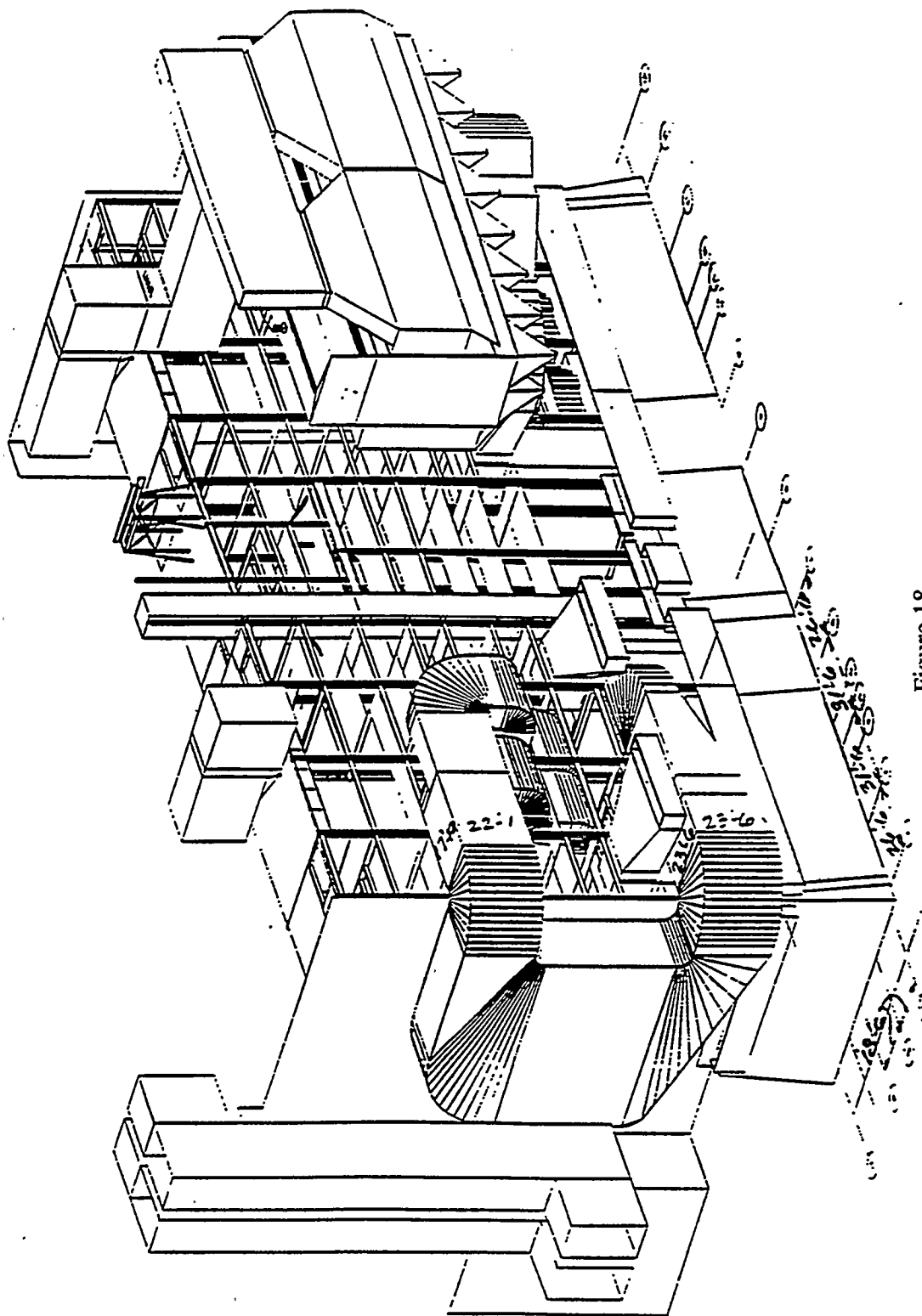
Figure 18
Perspective View of Retrofit SCR Arrangement for Plant D
(Electrostatic Precipitator (ESP) Removed for Clarity)
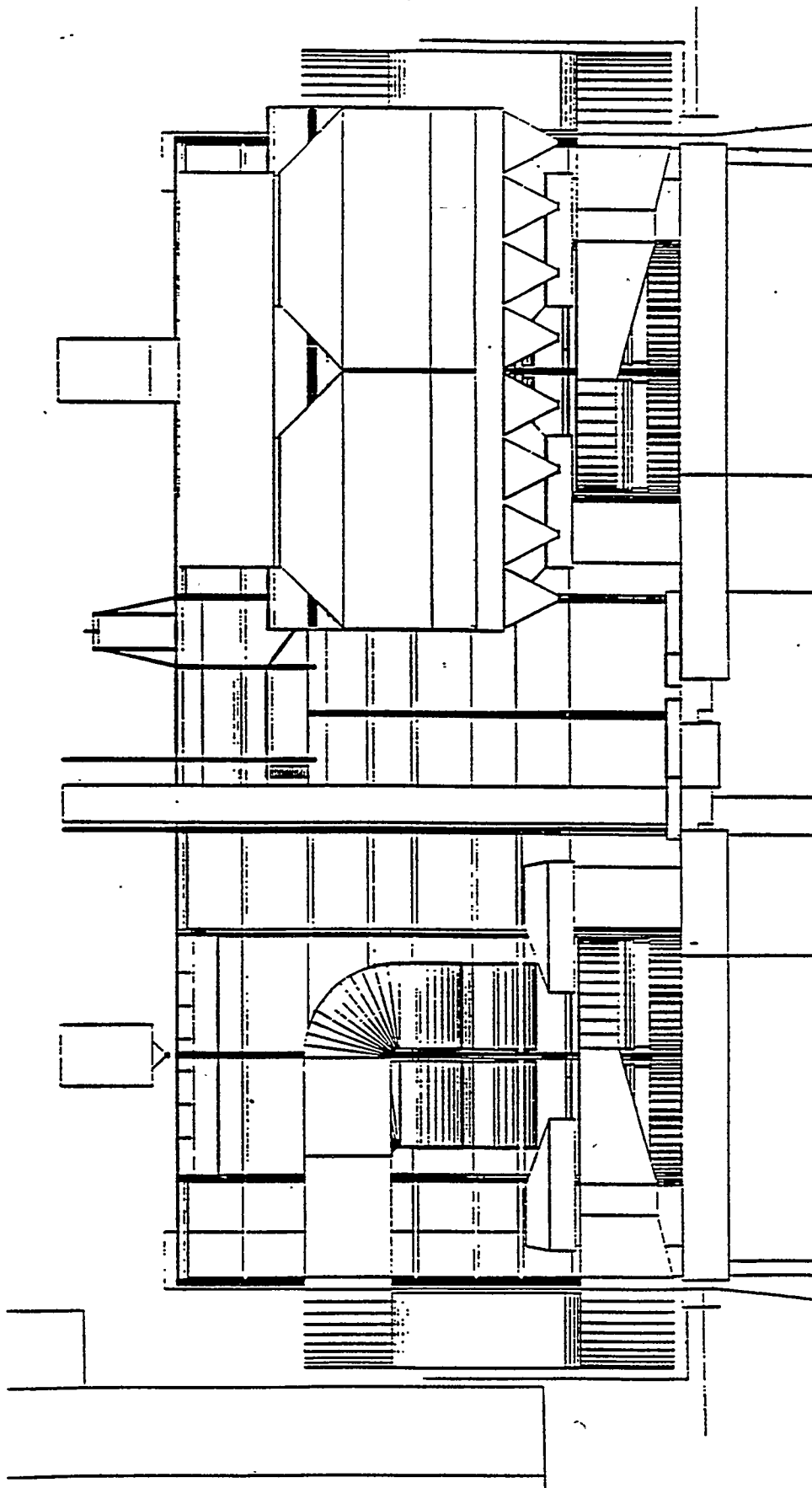
51

Figure 19
Front Elevation View of Retrofit SCR Arrangement for Plant D
(Electrostatic Precipitator (ESP) Removed for Clarity)

configuration, which places the reactors on opposite sides of the boiler house. The economizer outlet duct would turn up and exit the boiler room wall above elevation 831 ft. Due to the very restricted space between the boiler house wall and the precipitator, approximately 19 ft, a vertically oriented narrow duct is proposed which would combine the two flue gas ducts to a single SCR reactor. While the reactor, as shown in the figures, is wrapped around the boiler building, this is not a requirement, and could project parallel to the existing railroad track beneath the precipitators. The SCR exit duct would reenter the boiler house below elevation 820 ft, again having to fit through a very restricted space between the boiler house wall and the precipitator. As seen in the figures, this return duct is likely to block access to the existing road located under the precipitator ductwork.

The most difficult aspect of SCR retrofit at these two units is the tie-in points at the economizer outlet and air preheater inlet. The conceptual arrangement requires several 90-degree turns in the SCR inlet and outlet ductwork to get out of the boiler house. It is thought that an alternate, more optimized, ductwork routing could be achieved with a more detailed study. One alternative may allow routing of the SCR inlet duct inside the boiler house, running parallel to the back wall. This would allow penetration of the SCR inlet duct on the side of the existing boiler house wall. However, this would block several bays inside the boiler house. The SCR return duct would require that the existing flue gas conditioning equipment be removed to make room for the new ductwork.

While ID fan allowances are included in the estimate, these units currently utilize retrofitted two-speed ID fans. Normal practice is to run the ID fan on lower speeds most of the time. Therefore, it may be possible to accommodate the additional SCR draft loss with minimal modifications to the existing ID fan.

4.2.5   Plant E

Plant E includes three tangentially fired, subcritical units nominally rated at 100 MW and two tangentially fired, subcritial units nominally rated at 125 MW. Each boiler has four elevations of coal nozzles in cast iron windboxes located in each of the four corners. The boilers are fired under balanced pressure. None of the 100-MW units have been retrofitted with low-NO_x firing systems. The two 125-MW units have been retrofitted with low-NO_x firing systems featuring flame attachment nozzles, offset secondary air, and two elevations of close coupled overfire air.

The five Plant E units are similar to the two Plant D units in that the electrostatic precipitator effectively blocks access to the tie-in point between the boiler economizer exit and the air preheater inlet. This is exacerbated by the fact that the air preheaters are located below grade elevation, while the precipitators are located above grade elevation, blocking well over half of the back side of the boiler building. The precipitators are also located very close to the boiler building, which complicates the ability to pass a duct between the precipitator and the boiler house. Since there are multiple adjacent units at Plant E (units arranged in a side-by-side power block), access to the interior units will complicate the retrofit. Space is available on the boiler house roof to accommodate an SCR reactor having preliminary dimensions of 530 square ft cross-sectional area by 30 ft depth for each unit.

The conceptual design includes rerouting the existing duct from the economizer to the air preheater by replacing the 90-degree elbow down to the air preheater with a 90-degree elbow up toward the roof. The supply duct from the economizer outlet to the SCR would run from the economizer inside the boiler house, to an approximate elevation of 787 ft, until the precipitator located outside the boiler house wall is cleared. The duct would then be run outside the boiler house up to the SCR. The maximum dimension for the supply duct, while inside the boiler house, would be 18 ft 11 in. by 3 ft 10 in. The duct size, once outside the boiler house wall, is essentially unrestricted.

The return duct will run posterior to the supply duct and enter the boiler house wall below the supply duct, then elbow down to the existing transition piece on the air preheater. The return duct from the SCR to the air preheater can be routed outside the boiler house wall to approximate elevation 785 ft, at which point the duct can penetrate the boiler house wall and run inside the boiler house to the air preheater. The maximum dimension for the return duct, once inside the boiler house, is 18 ft 11 in. by 3 ft 8 in. The duct located outside the boiler house wall is unrestricted. Preliminary observations indicate that the transition duct from the air preheater to the new duct (on all units) could remain; however, turning vanes most likely will be required.

Units at Plant E, under the configuration described above, will require some major structural steel modifications to accommodate the duct run. Additionally, there are several large diameter ash pipes and roof drain pipes that will require relocation.

Additional retrofit difficulty also exists because all five units share a common chimney. Flue gas ductwork from the 100-MW units are combined, entering on one side of the stack. Ductwork from the 125-MW units are combined, entering on the opposite side of the stack. Because the units share a common stack there is little potential for rearranging or relocating components to make room for the SCR related equipment behind the boiler house.

4.2.6 Plant F

Plant F includes two tangentially fired, subcritical units nominally rated at 350 MW each. There are five elevations of coal nozzles in each of the four corners. The boiler is fired under positive pressure. Both units are designed to deliver superheated steam at a rate of 2,568,331 lb/hr at turbine inlet conditions of 1000°F and 2400 psig. Both units have been retrofitted with low-NO_x firing systems featuring flame attachment coal nozzles, offset secondary air, and SOFA. There are two SOFA boxes, with two compartments each, on each of the four corners. Each compartment contains tilting air nozzles and individual damper control.

The arrangement of the electrostatic precipitators at Unit F would allow access through the back of the boiler house wall to the boiler economizer exit and air preheater inlet. Two locations were identified for the SCR reactor.

The first location indicates that the boiler house roof would have sufficient space for the SCR reactor. Alternatively, the SCR reactors could be located on structural towers over the existing electrostatic precipitator inlet ductwork and the precipitator itself.

The supply duct could be routed by extending the existing elbow at the economizer to outside the boiler house and turning the elbow up to run the duct to the SCR. This would require the use of turning vanes because of the restrictive height between the economizer and the air preheater of less than 20 ft.

The return duct from the SCR to the air preheater could be routed outside the boiler house, posterior to the supply duct, and then penetrate inside at an elevation of approximately 835 ft. The transition piece from the air preheater to the duct run probably would not need to be adjusted. Turning vanes would be required to allow for the restrictive space to turn into the air preheater transition duct. The dimension of the duct outside the boiler house is unrestricted. Some structural steel modification would be required to accommodate the revised duct for both the supply and return.

In addition to moderate retrofit difficulty due to ductwork and reactor location, a balanced draft conversion on both units would likely be required in order to accommodate the increased draft loss due to the SCR. The addition of ID fans and balanced draft conversion are reflected in the capital cost and increase considerably the cost of adding SCR to these units.

4.3 Cost Methodology

Retrofitting SCR to an existing plant requires higher capital cost than a new plant because of the need to integrate the process into existing plant systems and accommodate site-specific physical and operational constraints. In addition, when compared to new boilers, higher inlet NO_x from existing boilers will necessitate greater catalyst volumes and, therefore, larger reactor sizes in an application where there is likely to be less space available due to retrofit difficulties. When compared with new installations, necessary costs for upgrade or new ID fans, gas handling equipment, and balance-of-plant modifications are often not included in literature estimates of SCR retrofit cost. These costs are included in this analysis.

This section describes the economic considerations and methods used to evaluate SCR as a potential retrofit NO_x control technology for selected units included in this study.

4.3.1 Economic/Technical Assumptions

The following technical and economic assumptions were used in this retrofit study:

- The retrofit study considered tangenially fired units with boiler sizes ranging from 100 MW to 955 MW.

- SCR design removal efficiencies of 40 percent, 60 percent, and 80 percent were estimated.

- Catalyst life guarantee was assumed to be 2 years (16,000 hours).

- Ammonia slip was assumed to be 2 ppmv measured on a dry basis.

- Actual inlet NO_x concentrations for the existing units were used as the basis of the SCR design. The inlet NO_x ranged from 0.55 to 0.40 lb/MBtu and represent tangentially fired units both with and without low NO_x combustion modifications.

- Similar to a new unit, it was assumed that the required operation of the SCR was over a boiler load range of 35 percent to 100 percent.

- An SO_2 to SO_3 oxidation rate of 1.0 percent was assumed due to a lower sulfur coal. (In the case of the new unit analysis utilizing a nominal 3 percent sulfur fuel, the lowest possible oxidation rate of 0.75 percent was desired to minimize the collateral impacts of high SO_3 concentrations. However, in the case of the retrofit analysis where a nominal 1.5 percent sulfur fuel is used, it was thought that a slightly higher oxidation rate of 1 percent could be tolerated in an effort to maximize space velocity for a given NO_x reduction, resulting in an overall reduction in SO_3 concentration at the SCR outlet compared to the new unit case.)

- A 15 year life was assumed. Units which were currently scheduled to retire prior to the end of the study are assumed to be extended through the end of the study period.

- All costs are expressed in 1996 dollars.

- Heat input to each boiler is the 15-year average annual total Btu projected burn for the unit before the SCR was added.

- The velocity, ammonia, and temperature distribution requirements are assumed to be identical to the new unit analysis shown in table 3.

  - The NO_x rate for each unit is assumed to be the rate at the unit's operating maximum.

  - The NO_x rates are assumed not to affect the economic dispatch of the units.

- Eastern low-sulfur coal (nominally 1 to 1.5 percent sulfur) was assumed to be the fuel for all units.

- Increases in station service and/or heat rate impacts are valued using SCS's Worth of Unit Improvement (WUI) methodology. The WUI is a methodology for valuing the additional station service consumed and the heat rate impacts due to the addition of a particular NO_x control technology. Calculations are specific to the Southern electric system and take into account each unit's total fuel cost, O&M cost, unit capacity factors, and hours of operation at various output levels for each unit. Therefore, depending on the particular unit, the value of station service will be greatest at lower loads when the unit is less efficient, and smallest at higher loads when the unit is operating at its maximum efficiency. The WUI method also considers the value impact to the system due to changes in system hourly production cost and capacity deferment.

56

- The reactor assumed for each application in this study utilizes a hot-side, high-dust configuration with three catalyst layers plus a flow straightener layer. The flow straightener layer consists of fabricated modules of 2 in. x 2 in., 16-gauge mild steel tube approximately 18 inches in length.

- The design includes one vertical, downflow reactor per air preheater. Therefore, on units where a split train draft system utilizes two air preheaters, two SCR reactors are included in the estimate.

- The reactor is equipped with an economizer bypass to permit SCR operation at lower boiler loads.

- All catalyst layers include steam sootblowers. The sootblower design is identical to those used in the SCR demonstration project facility.

4.3.2 Capital Costs

Capital costs for the SCR include all ammonia storage and injection equipment, reactor with initial catalyst charges, allowance for ID fan upgrade (or balanced draft conversion cost), allowance for air preheater upgrade, erection, indirects, AFUDC, engineering, temporary construction facilities, utility company overheads, and field supervision.

4.3.3 O&M Costs

Fixed O&M costs include estimates of maintenance material and labor, operating labor, and administration/support labor. Variable O&M captures ammonia consumption and catalyst replacement costs. In addition, estimates of incremental station service costs due to SCR and minimum SCR load point for calculating thermal efficiency (heat rate) penalty are included.

4.4 Summary of Capital and O&M Costs for Each Unit

Table 16 shows the capital, O&M, and current dollar levelized costs for selected units. The SCR retrofit costs vary from $1,541/ton to $7,419/ton depending on NO_x removal percentage, unit size, inlet NO_x concentration, utilization (capacity factor), and capital and O&M costs. Even though the capital cost (in dollars) increases as plant size increases, lower levelized costs are achieved when SCR is applied to larger, higher utilized units such as Plants A, B, and C. This is due to economies of scale and the fact that the quantities of NO_x removed are greater on larger units.

All of the units shown in table 16 have been retrofitted with some type of combustion modifications to lower the NO_x concentration prior to the retrofit of SCR. While some capital cost savings can be achieved in the SCR by lowering the inlet NO_x, the resulting levelized cost is higher due to fewer tons removed when compared to an SCR retrofit on an uncontrolled unit.

57

Table 16
Capital, O&M, and Levelized Cost for Retrofit of SCR to Selected Southern Company Units

Power Plant Attributes	Units	Plant A	Plant B	Plant C	Plant D	Plant E	Plant F
Plant Capacity	MW	700	880	880	265	100	350
Average Annual Heat Input	MBtu	43,137,998	57,714,319	54,107,012	9,848,867	3,627,034	17,746,062
Calculated Capacity Factor	%	74%	81%	75%	43%	33%	55%
Evaluation life	years	15	15	15	15	15	15
40 Percent Removal							
SCR Removal Efficiency	%	40	40	40	40	40	40
Emission without SCR	lb/MBtu	0.41	0.43	0.41	0.37	0.45	0.32
Emission with SCR	lb/MBtu	0.25	0.28	0.25	0.22	0.27	0.19
Tons of NOx removed	ton/yr	3537	4963	4437	729	328	1138
CAPITAL COST							
Capital Cost ($) *(note 1)*		$41,933,000	$110,795,000	$48,131,000	$19,778,000	$8,112,000	$37,639,000
Capital Cost ($/kw)		$60	$126	$55	$75	$81	$108
ANNUAL OPERATING AND MAINTENANCE COST							
Fixed and Variable Operating Costs		$2,062,000	$2,453,000	$2,453,000	$1,116,000	$757,000	$1,300,000
WUI Operating Costs *(note 4)*		$601,000	$691,000	$708,000	$304,000	$257,000	$767,000
LEVELIZED COST							
Current Dollar Levelized Cost	$/ton	$2,700	$4,181	$2,498	$6,440	$7,419	$7,220
60 Percent Removal							
SCR Removal Efficiency	%	60	60	60	60	60	60
Emission without SCR	lb/MBtu	0.41	0.43	0.41	0.37	0.45	0.32
Emission with SCR	lb/MBtu	0.16	0.17	0.16	0.15	0.18	0.13
Tons of NOx removed	ton/yr	5306	7445	6655	1093	490	1704
CAPITAL COST							
Capital Cost ($) *(note 1)*		$45,295,000	$114,440,000	$52,019,000	$21,336,000	$8,747,000	$39,188,000
Capital Cost ($/kw)		$65	$130	$59	$81	$87	$112
ANNUAL OPERATING AND MAINTENANCE COST							
Fixed and Variable Operating Costs		$2,461,000	$2,949,000	$2,949,000	$1,283,000	$836,000	$1,513,000
WUI Operating Costs *(note 4)*		$630,000	$732,000	$744,000	$317,000	$283,000	$782,000
LEVELIZED COST							
Current Dollar Levelized Cost	$/ton	$1,991	$2,946	$1,848	$4,703	$5,346	$5,108
80 Percent Removal							
SCR Removal Efficiency	%	85	80	80	80	80	80
Emission without SCR	lb/MBtu	0.41	0.43	0.41	0.37	0.45	0.32
Emission with SCR	lb/MBtu	0.06	0.09	0.08	0.07	0.09	0.06
Tons of NOx removed	ton/yr	7517	9927	8874	1458	653	2271
CAPITAL COST							
Capital Cost ($) *(note 1)*		$47,935,000	$116,030,000	$55,112,000	$22,519,000	$9,223,000	$39,762,000
Capital Cost ($/kw)		$68	$132	$63	$85	$92	$114
ANNUAL OPERATING AND MAINTENANCE COST							
Fixed and Variable Operating Costs		$3,099,000	$3,714,000	$3,714,000	$1,613,000	$1,049,000	$1,903,000
WUI Operating Costs *(note 4)*		$641,000	$746,000	$759,000	$322,000	$265,000	$788,000
LEVELIZED COST							
Current Dollar Levelized Cost	$/ton	$1,559	$2,325	$1,541	$3,917	$4,502	$4,071

Notes:
1. Capital cost estimate includes the cost of balance draft conversion for Plant B and Plant F.
2. Levelized cost based on 15 year life, 9.245% cost of capital, and 3.04% escalation.
3. All values shown in 1996 dollars.
4. Worth of Unit Improvement (WUI) methodology is used to value the heat rate impacts and additional station service requirements.

Figure 20 shows a comparison of levelized cost vs. NO_x removal efficiency for a new and retrofit SCR installation applied to a 250-MW unit designed for 60 percent removal. While the retrofit unit levelized cost is higher than the new unit, the difference is fairly small. The difference is primarily due to higher capital cost of the retrofit installation, since the inlet NO_x concentrations for the retrofit (0.40 lb/MBtu) and the new unit (0.35 lb/MBtu) are similar and approximately the same number of tons of NO_x are removed.

While figure 20 shows little difference in levelized costs, the capital cost differences between the new SCR installation and the retrofit SCR installations are large. For a 60 percent removal design, the estimated retrofit cost is approximately 51 percent greater than the estimated new cost installation. (This comparison is highly site specific and actual retrofit costs may be higher or lower than those presented here) Table 17 shows the capital cost difference between a new and retrofit SCR installation.

Table 17
Capital Cost Differences for New and Retrofit SCR Installations
(250-MW Plant Size)

	NO_x Removal Efficiency		
	40%	**60%**	**80%**
New SCR Installation			
Total Capital Requirement	$12,974,000	$13,415,000	$14,142,000
Total Capital Requirement	$52/kW	$54/kW	$57/kW
Retrofit SCR Installation			
Total Capital Requirement	$18,800,000	$20,281,000	$21,403,000
Total Capital Requirement	$75/kW	$81/kW	$86/kW

As seen from table 17 and figure 20, technical and economic assessment of SCR must be based on both the cost effectiveness and the first cost (capital cost cash flow) of the proposed installation.

4.5   Extrapolation of Data to High Inlet NO_x Cases

At the request of the DOE, the Southern Company specific retrofit cost data were extrapolated to high inlet NO_x conditions in an effort to represent many of the boilers in the OTAG region.

Due to the fact that Southern Company's boiler population is predominantly wall-fired and tangentially fired boilers, there was has been no corporate need to rigorously define the capital and operating cost at high inlet NO_x concentrations indicative of cell burner and cyclone fired boilers. However, it is recognized that these high emitting boiler types may face more stringent

Levelized Cost vs. NOx Removal
SCS/DOE Project - Economic Evaluation
Based on 250 MW Unit and 60% Removal

Retrofit SCR (Inlet NOx = .40 lb/MBtu)

New SCR (Inlet NOx = .35 lb/MBtu)
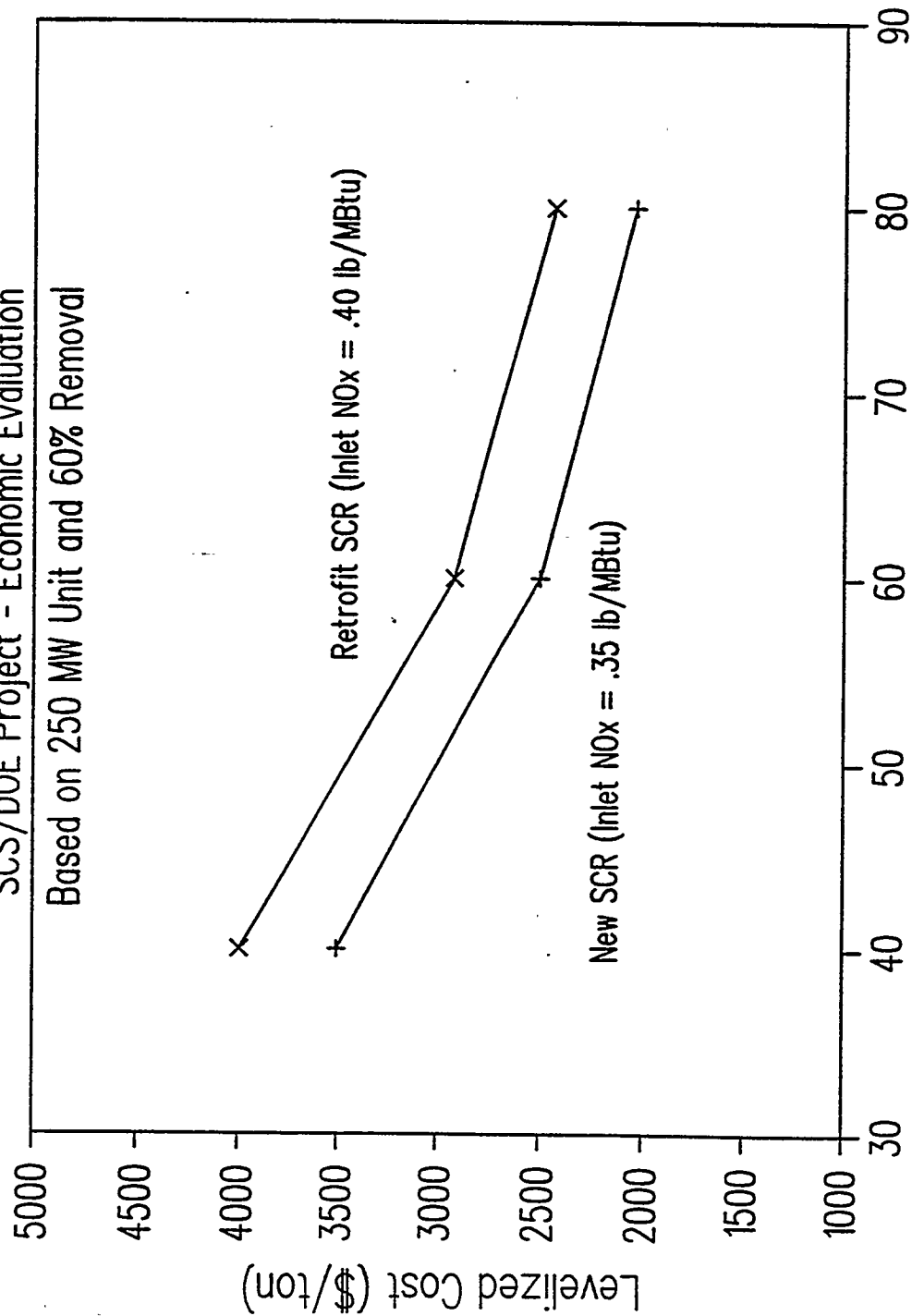
NOx Removal (%)
Figure 20

Levelized Cost ($/ton)

RETREM.atb

60

NO$_x$ control requirements in the future. Therefore, analysis of high emitting boilers is presented in this report for information purposes.

The results presented in this section are subject to the following caveats:

1. The estimate is based on a 250-MW unit with a retrofitted SCR designed for 60 percent removal. The retrofit difficulty is representative of plant configurations in the OTAG region.

2. Best efforts were made to adjust the capital and O&M costs for increasing inlet NO$_x$ conditions. Specifically, the space velocity (catalyst volume), reactor height, and ammonia consumption are the primary process variables adjusted. These adjustments were based on factors obtained from several catalyst suppliers.

3. This comparison, while valid for screening purposes, is generic in nature and does not preclude the need to perform site specific cost evaluations, particularly for high NO$_x$ emitting boilers.

The results of the extrapolated data shown in figure 21 indicate a significant trend of decreasing levelized cost with increasing inlet NO$_x$ concentration, highlighting a key difference in levelized cost between lower emitting boilers (or boilers which have been controlled with combustion modifications prior to the SCR) and higher emitting, uncontrolled boilers.

For a given removal percentage, 60 percent in this case, the higher emitting boilers result in a greater number of tons removed when compared to the lower emitting boilers. The annual tons of NO$_x$ removed range from 1247 to 9975 for 0.30 lb/MBtu and 2.4 lb/MBtu inlet NO$_x$ concentrations, respectively.

Note that the constant dollar levelized cost of $534/ton at 2.4 lb/MBtu inlet NO$_x$ is reasonably consistent with the levelized cost of $404/ton for a 65 percent NO$_x$ reduction system recently reported by Public Service of New Hampshire at their Merrimack station.[16] (Differences between the two values are mainly attributable to the study retrofit cost of $87/kW vs. Merrimack's recently reported capital cost of $56/kW, and the difference between in 60 percent and 65 percent removal.) Results shown in figure 21 also compare favorably with recently published levelized cost from Pubic Service Electric and Gas Company's Mercer Generating Station. [19]

Levelized Cost vs. High Inlet NOx
SCS/DOE Project - Economic Evaluation

Based on Retrofit 250 MW Unit and 60% Removal
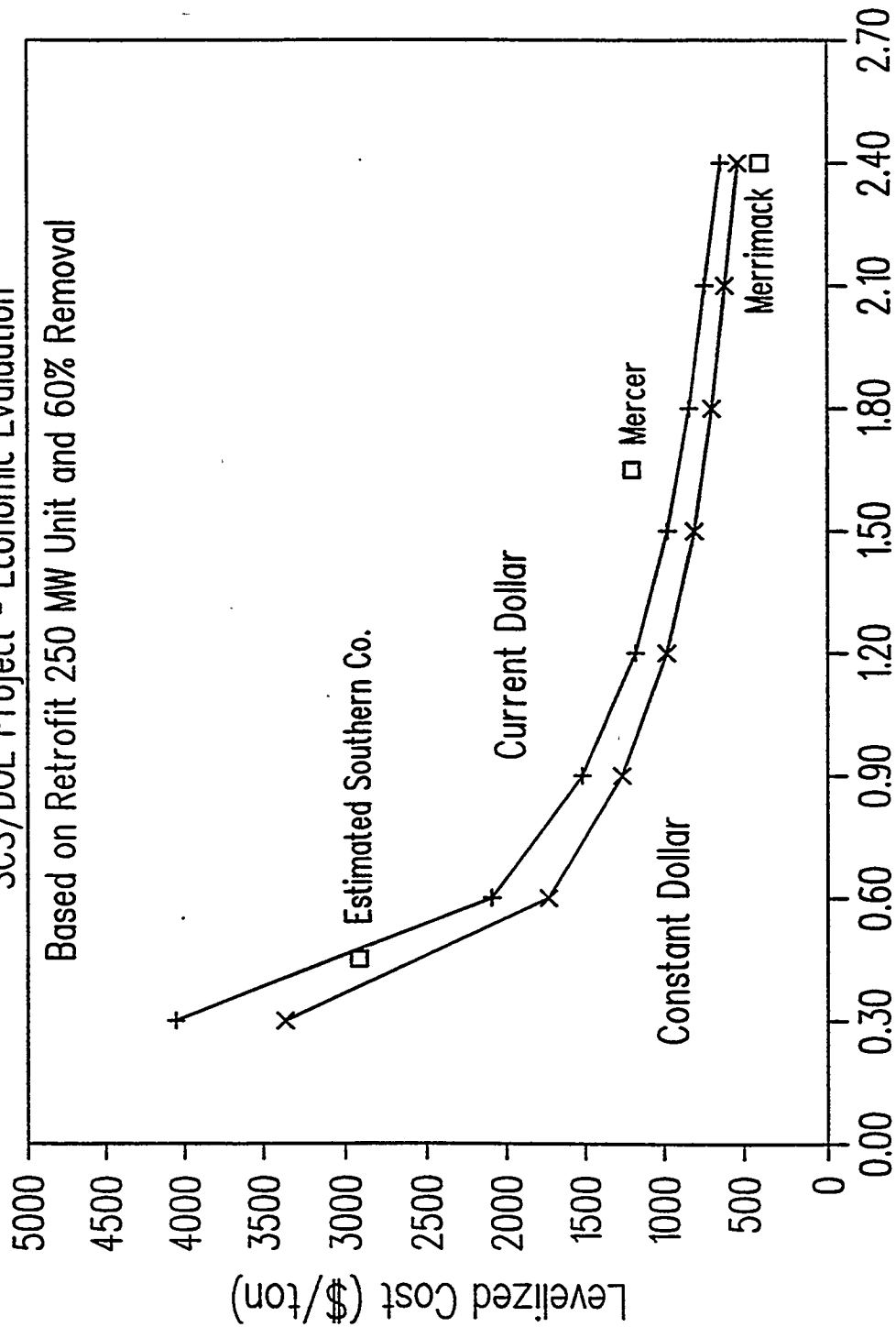
Figure 21

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62

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Exhibit A

Economic Evaluation Parameters

EXHIBIT A
ECONOMIC EVALUATION PARAMETERS

PROJECT TITLE: DOE SCR Project - Economic Evaluation

Economic Assumptions	Units	Value
Cost of debt	%	8.5
Dividend rate for preferred stock (pre-tax)	%	7.0
Dividend rate for common stock (pre-tax)	%	11.0
Debt ratio, % of total capital		50.0
Preferred stock, % of total capital		15.0
Common stock, % of total capital		35.0
Income tax rate	%	38.0
Investment tax credit	%	0.0
Property taxes & insurance	%	3.0
Inflation rate	%	3.0
Discount rate (with inflation)	%	9.150
Discount rate (without inflation)	%	5.971
Escalation of raw materials above inflation	%	0.0
Construction period (choose 1-6)	Years	1.5
Allowance for funds during construction [a]	%	1.91
Construction downtime	Days	0
Remaining life of power plant	Years	30
Year for input cost data		1996
Year for costs presented in this report		1996
Royalty allowance (based on Total Process Capital)	%	0
Capital charge factor & O&M cost levelization factor		
Current dollars:		
Capital charge factor		0.150
O & M cost levelization factor		1.362
Constant dollars:		
Capital charge factor		0.116
O & M cost levelization factor		1.000
Power plant capacity factor	%	65
Sales tax rate	%	5.0
Cost of freight for process equipment	%	2.0
Sales tax plus freight	%	7.0
General facilities, % of total process capital		2.0
Engineering & home office fees, % of total process capital		8.0

---

[a] Represents the time value of funds used for construction based on an interest rate equal to weighted cost of capital assuming 3.0 % inflation rate and 9.150 % discount rate.

Calculation of Capital Charge Factor and O&M Cost Levelization Factor

	Units	Value	Equivalent Value	Without Inflation
Cost of debt	%	8.5	0.085	0.053
Dividend rate for preferred stock (pre-t)	%	7	0.070	
Dividend rate for common stock (pre-t)	%	11	0.110	
Debt ratio, % of total capital		50.0	0.50	
Preferred stock, % of total capital		15.0	0.15	
Equity Ratio, % of total capital		50.0	0.50	
Income tax rate	%	38.0	0.38	
Investment tax credit (ITC)	%	0.0	0.00	
Property taxes & insurance	%	3.0	0.03	
Inflation rate	%	3.0	0.03	
Discount rate (with inflation)	%	9.15		
Discount rate (without inflation)	%	5.971		
Remaining life of power plant	Years	30		
Tax recovery period	Years	30		
Straight line tax depreciation	%/yr	3.333	0.0333	
Equity cost	%	9.800	0.0080	0.0060
Book depreciation rate net of ITC	%/yr	3.333	0.0333	
Book value net of ITC	%	100	1.00	

m = tax recovery life, years
n = book life, years

Current Dollars (With Inflation)

Columns: Year f, Present Value Factor v, Annuity Factor a, Annuity Depreciation a, Tax Depreciation Rate td, Deferred Income Tax dt, Taxes Paid tp, Year-by-Year Carrying Charge CC, Present Value of Carrying Charge, Cumulative Present Value CC Vf,m,n, Levelized CC Pf,m,n

Constant Dollars (Without Inflation)

Columns: Year f, Fractional Remaining Book Value SB, Return on Equity RE, Return on Debt RD, Present Value Factor pv, Annuity Factor a, Taxes Paid tp, Year-by-Year Carrying Charge CC, Present Value of Carrying Charge CC, Cumulative Present Value CC Vf,m,n, Levelized CC Pf,m,n, Return on Equity RE, Return on Debt RD

O&M Cost Levelization Factor Calculations

	Units	Current $	Constant $
Inflation rate	%	3.00	0.00
Discount rate	%	9.15	5.971
Escalation of raw materials above inflation	%	0.00	0.00
Remaining life of power plant	Years	30	30
Apparent escalation rate	%	3.00	0.00
Escalation factor		0.943858	0.943858
Annuity factor		10.138528	13.807837
O&M cost levelization factor		1.361918	1.000000

Exhibit B

250-MW New Plant Combustion Calculation

filename DOE250.wk4
DOE SCR Project - Economic Evaluation

Name.: ECH
Date: 3/22/96 (rev 1), 4/1/96 (rev2), 8/29/95 (rev 3)
Project: DOE SCR Project - Base Case Unit (250 MW) Combustion Calculation

Input Data

Coal Source	Typical High Sulfur
Heating Value (Btu/lb)	12500
Plant Heat Rate (BTU/kwh)	9500
Combustion Air Moisture (#H2O/#Dry Air)	0.013
Calculated Excess Air (%)	18
Unit Load (MW)	250
Flue Gas Temp (F)	700
Flue Gas Pressure (In. W G)	-5 14.52 (psia)

Coal Composition	Weight Percent
C	67.48
H	4.51
N	1.43
S	2.33
Cl	0.14
O	5.92
H2O	8.39
ash	9.80
Total	100.00

Field Measured Values	
Measured O2 (% wet)	3.00
Measured SO3 (ppm wet)	0
Measured NOx (ppm wet)	230
Measured Particulate (high) (mg/Nm3)	0
Measured Particulate (low) (mg/Nm3)	0

Combustion Calculation Output Data

Combustion Products	Flue Gas Flow Rate (#mol/h)	Flue Gas Comp (mol%)	Flue Gas Flow Rate (#/h)	Flue Gas Comp (wt%)	Flue Gas Flow Rate (scfm)	Flue Gas Flow Rate (acfm)
CO2	10675.44	14.067	469826	20.775	63875	144286
O2	2284.26	3.010	73096	3.232	13668	30873
N2	56166.33	74.013	1573219	69.566	336062	759125
SO2	138.08	0.1820	8846	0.391	826	1866
SO3	1.38	0.00182	111	0.00489	8	18
NO	16.58	0.02185	497	0.02200	99	224.11
NO2	0.87	0.00115	40	0.00178	5	11.80
HCl	4.50	0.00593	164	0.00726	27	60.84
H2O	6600.13	8.697	118934	5.259	39491	89205
ash			16758	0.741		
Total	75,888	100.000	2,261,492	100.000	454,061	1,025,671

Summary	
Calculated O2 (% wet):	3.01
Calculated O2 (% dry):	3.30
Calculated SO3 (ppm wet):	18
Calculated NOx (ppm wet):	230
Calculated NOx (ppm wet corr):	230
Calculated NOx (ppm dry):	252
Calculated NOx (ppm dry corr):	256
Calculated NOx (lb/MBTU):	0.35
Calculated Coal Feed (lb/hr):	190,000
Calculated Heat Input (MBTU/hr):	2,375

Exhibit C

250-MW New Plant - SCR Capital, O&M, and Levelized Costs for 40% Removal

Exhibit C

250 MW Plant - SCR Capital Cost for 40% Removal

Process Areas	k$	$/kw
Catalyst	$1,856	$7.4
Reactor Housing, Ductwork, Steel	$4,958	$19.8
Sootblowers	$580	$2.3
Ammonia Storage, Handling, and Injection	$1,292	$5.2
ID Fan Differential	$216	$0.9
Air Preheater Differential	$220	$0.9
Ash Handling Differential	$300	$1.2
Electrical	$201	$0.8
Instruments & Controls	$100	$0.4
Testing, Training, Commissioning	$138	$0.6
(A) Total Process Capital (sum of process areas)	$9,861	$39.4
(B) General Facilities (2% of A)	$197	$0.8
(C) Engineering (8% of A)	$789	$3.2
(D) Project Contingency (15% of A+B+C)	$1,627	$6.5
(E) Total Plant Cost (A+B+C+D)	$12,474	$49.9
(F) Allowance for Funds During Construction (1.91% of E)	$238	$1.0
(G) Total Plant Investment (E+F)	$12,712	$50.8
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$159	$0.6
(J) Inventory Capital	$103	$0.4
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$12,974	$51.9

Exhibit C

250 MW Plant - SCR Operating and Maintenance Cost for 40% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$79,000
Maintenance material				$118,000
Administration/support labor				$43,000
Subtotal Fixed Costs				$305,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	3.56	$2.00	$41,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	125	$0.13	$89,000
Catalyst	cu. ft.	(Note 1)	$400	$385,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	622	$0.03	$106,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$621,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$926,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or replacement schedule. Dollar amount shown in this table represents a levelized annual reserve
for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit C

250 MW Plant - Summary of Performance and Cost Data for 40% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	40
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.21
Amount removed	ton/yr	916

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.454	0.116	1.125
Fixed O&M Cost	1.362	0.311	1.000	0.228
Variable O&M Cost	1.362	0.632	1.000	0.384
Total Cost		2.397		1.737
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$2,124	0.116	$1,642
Fixed O&M Cost	1.362	$454	1.000	$333
Variable O&M Cost	1.362	$924	1.000	$561
Total Cost		$3,502		$2,536

Exhibit D

**250-MW New Plant - SCR Capital, O&M, and Levelized Costs for 60% Removal
(Base Case)**

Exhibit D

250 MW Base Case SCR Capital Cost for 60% Removal

Process Areas	k$	$/kw
Catalyst	$2,168	$8.7
Reactor Housing, Ductwork, Steel	$4,958	$19.8
Sootblowers	$580	$2.3
Ammonia Storage, Handling, and Injection	$1,292	$5.2
ID Fan Differential	$216	$0.9
Air Preheater Differential	$220	$0.9
Ash Handling Differential	$300	$1.2
Electrical	$201	$0.8
Instruments & Controls	$100	$0.4
Testing, Training, Commissioning	$138	$0.6
(A) Total Process Capital (sum of process areas)	$10,172	$40.7
(B) General Facilities (2% of A)	$203	$0.8
(C) Engineering (8% of A)	$814	$3.3
(D) Project Contingency (15% of A+B+C)	$1,678	$6.7
(E) Total Plant Cost (A+B+C+D)	$12,868	$51.5
(F) Allowance for Funds During Construction (1.91% of E)	$246	$1.0
(G) Total Plant Investment (E+F)	$13,114	$52.5
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$179	$0.7
(J) Inventory Capital	$122	$0.5
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$13,415	$53.7

Exhibit D

250 MW Base Case SCR Operating and Maintenance Cost for 60% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$81,000
Maintenance material				$122,000
Administration/support labor				$44,000
Subtotal Fixed Costs				$312,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	3.56	$2.00	$41,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	187	$0.13	$133,000
Catalyst	cu. ft.	(Note 1)	$400	$450,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	639	$0.03	$109,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$733,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$1,045,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or
replacement schedule. Dollar amount shown in this table represents a levelized annual reserve
for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit D

250 MW Base Case Summary of Performance and Cost Data for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.746	1.000	0.454
Total Cost		2.569		1.851
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$726	1.000	$442
Total Cost		$2,500		$1,802

Exhibit E

250-MW New Plant - SCR Capital, O&M, and Levelized Costs for 80% Removal

Exhibit E

250 MW Plant - SCR Capital Cost for 80% Removal

Process Areas	k$	$/kw
Catalyst	$2,536	$10.1
Reactor Housing, Ductwork, Steel	$4,958	$19.8
Sootblowers	$580	$2.3
Ammonia Storage, Handling, and Injection	$1,454	$5.8
ID Fan Differential	$216	$0.9
Air Preheater Differential	$220	$0.9
Ash Handling Differential	$300	$1.2
Electrical	$201	$0.8
Instruments & Controls	$100	$0.4
Testing, Training, Commissioning	$138	$0.6
(A) Total Process Capital (sum of process areas)	$10,702	$42.8
(B) General Facilities (2% of A)	$214	$0.9
(C) Engineering (8% of A)	$856	$3.4
(D) Project Contingency (15% of A+B+C)	$1,766	$7.1
(E) Total Plant Cost (A+B+C+D)	$13,538	$54.2
(F) Allowance for Funds During Construction (1.91% of E)	$259	$1.0
(G) Total Plant Investment (E+F)	$13,797	$55.2
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$202	$0.8
(J) Inventory Capital	$143	$0.6
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$14,142	$56.6

Exhibit E

250 MW Plant - SCR Operating and Maintenance Cost for 80% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$86,000
Maintenance material				$128,000
Administration/support labor				$45,000
Subtotal Fixed Costs				$324,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	3.56	$2.00	$41,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	250	$0.13	$178,000
Catalyst	cu. ft.	(Note 1)	$400	$526,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	655	$0.03	$112,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$857,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$1,181,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or
replacement schedule. Dollar amount shown in this table represents a levelized annual reserve
for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit E

250 MW Plant – Summary of Performance and Cost Data for 80% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	80
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.07
Amount removed	ton/yr	1833

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.585	0.116	1.226
Fixed O&M Cost	1.362	0.331	1.000	0.243
Variable O&M Cost	1.362	0.872	1.000	0.531
Total Cost		2.788		2.000
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,158	0.116	$895
Fixed O&M Cost	1.362	$241	1.000	$177
Variable O&M Cost	1.362	$637	1.000	$388
Total Cost		$2,036		$1,460

Exhibit F

125-MW New Plant - Combustion Calculation

DOE SCR Project - Economic Evaluation

		Name:	ECH
		Date:	3/22/96 (rev 1), 4/1/96 (rev2), 8/29/96 (rev 3)
		Project	DOE SCR Project - 125 MW Plant Combustion Calculation

Input Data

Coal Source	Typical High Sulfur		Coal Composition	Weight Percent		Field Measured Values	
Heating Value (Btu/lb)	12500		C	67.48		Measured O2 (% wet)	3.00
Plant Heat Rate (BTU/kwh)	9500		H	4.51		Measured SO3 (ppm wet)	0
Combustion Air Moisture (#H2O/#Dry Air)	0.013		N	1.43		Measured NOx (ppm wet)	230
			S	2.33		Measured Particulate (high) (mg/Nm3)	0
Calculated Excess Air (%)	18		Cl	0.14		Measured Particulate (low) (mg/Nm3)	0
Unit Load (MW)	125		O	5.92			
Flue Gas Temp (F)	700		H2O	8.39			
Flue Gas Pressure (In. W.G)	-5		ash	9.80			
	14.52 (psia)		Total	100.00			

Summary

Calculated O2 (% wet):	3.01
Calculated O2 (% dry):	3.30
Calculated SO3 (ppm wet):	18
Calculated NOx (ppm wet):	230
Calculated NOx (ppm wet corr):	230
Calculated NOx (ppm dry):	252
Calculated NOx (ppm dry corr):	256
Calculated NOx (lb/MBTU):	0.35
Calculated Coal Feed (lb/hr):	95,000
Calculated Heat Input (MBTU/hr):	1,188

Combustion Calculation Output Data

Combustion Products	Flue Gas Flow Rate (#mol/h)	Flue Gas Comp (mol%)	Flue Gas Flow Rate (#/h)	Flue Gas Comp (wt%)	Flue Gas Flow Rate (scfm)	Flue Gas Flow Rate (acfm)
CO2	5337.72	14.067	234913	20.775	31937	72143
O2	1142.13	3.010	36548	3.232	6834	15437
N2	28083.17	74.013	786609	69.566	168031	379563
SO2	69.04	0.1820	4423	0.391	933	933
SO3	0.69	0.00182	55	0.00489	4	9.33
NO	8.29	0.02185	249	0.02200	50	112.05
NO2	0.44	0.00115	20	0.00178	3	5.90
HCl	2.25	0.00593	82	0.00726	13	30.42
H2O	3300.07	8.697	59467	5.259	19745	44603
ash			8379	0.741		
Total	37,944	100.000	1,130,746	100.000	227,030	512,835

Exhibit G

125-MW New Plant - SCR Capital, O&M, and Levelized Costs for 60% Removal

Exhibit G

125 MW Plant - SCR Capital Cost for 60% Removal

Process Areas	k$	$/kw
Catalyst	$1,230	$9.8
Reactor Housing, Ductwork, Steel	$2,814	$22.5
Sootblowers	$329	$2.6
Ammonia Storage, Handling, and Injection	$733	$5.9
ID Fan Differential	$123	$1.0
Air Preheater Differential	$125	$1.0
Ash Handling Differential	$170	$1.4
Electrical	$114	$0.9
Instruments & Controls	$57	$0.5
Testing, Training, Commissioning	$78	$0.6
(A) Total Process Capital (sum of process areas)	$5,773	$46.2
(B) General Facilities (2% of A)	$115	$0.9
(C) Engineering (8% of A)	$462	$3.7
(D) Project Contingency (15% of A+B+C)	$952	$7.6
(E) Total Plant Cost (A+B+C+D)	$7,302	$58.4
(F) Allowance for Funds During Construction (1.91% of E)	$139	$1.1
(G) Total Plant Investment (E+F)	$7,442	$59.5
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$99	$0.8
(J) Inventory Capital	$61	$0.5
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$7.602	$60.8

Exhibit G

125 MW Plant - SCR Operating and Maintenance Cost for 60% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$46,000
Maintenance material				$69,000
Administration/support labor				$33,000
Subtotal Fixed Costs				$213,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	1.78	$2.00	$20,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	94	$0.13	$67,000
Catalyst	cu. ft.	(Note 1)	$400	$225,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	319	$0.03	$55,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$367,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$580,000

Note 1 – Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or replacement schedule. Dollar amount shown in this table represents a levelized annual reserve for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit G

125 MW Plant - Summary of Performance and Cost Data for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	117.5
Power produced, (net)	10^6 kWh/yr	669.045
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	270,465
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	687

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.704	0.116	1.318
Fixed O&M Cost	1.362	0.437	1.000	0.321
Variable O&M Cost	1.362	0.747	1.000	0.454
Total Cost		2.888		2.093
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,659	0.116	$1,283
Fixed O&M Cost	1.362	$425	1.000	$312
Variable O&M Cost	1.362	$727	1.000	$442
Total Cost		$2,811		$2,037

Exhibit H

700-MW New Plant - Combustion Calculation

filename: DOE700.wk4
DOE SCR Project - Economic Evaluation

Name: ECH
Date: 3/22/96 (rev 1), 4/1/96 (rev2), 8/29/96 (rev 3)
Project: DOE SCR Project - 700 MW Plant Combustion Calculation

Input Data

Coal Source	Typical High Sulfur	Coal Composition	Weight Percent		Field Measured Values	
Heating Value (Btu/lb)	12500	C	67.48		Measured O2 (% wel)	3.00
Plant Heat Rate (BTU/kwh)	9500	H	4.51		Measured SO3 (ppm wet)	0
Combustion Air Moisture (#H2O/#Dry Air)	0.013	N	1.43		Measured NOx (ppm wel)	230
		S	2.33		Measured Particulate (high) (mg/Nm3)	0
Calculated Excess Air (%)	18	Cl	0.14		Measured Particulate (low) (mg/Nm3)	0
		O	5.92			
Unit Load (MW)	700	H2O	8.39			
		ash	9.80			
Flue Gas Temp (F)	700	Total	100.00			
Flue Gas Pressure (in W.G)	-5		1452 (psia)			

Combustion Calculation Output Data

Combustion Products	Flue Gas Flow Rate (#mol/h)	Flue Gas Comp (mol%)	Flue Gas Flow Rate (#/h)	Flue Gas Comp (wt%)	Flue Gas Flow Rate (scfm)	Flue Gas Flow Rate (acfm)
CO2	29891.22	14.067	1315513	20.775	176849	404000
O2	6395.94	3.010	204670	3.232	38269	86445
N2	157265.72	74.013	4405013	69.568	940973	2125550
SO2	386.64	0.1820	24768	0.391	2313	5226
SO3	3.87	0.00182	310	0.00469	23	5226
NO	46.43	0.02185	1393	0.02200	278	627.51
NO2	2.44	0.00115	112	0.00178	15	33.03
HCl	12.60	0.00593	460	0.00728	75	170.35
H2O	18460.37	8.697	333016	6.259	110574	249774
ash			46922	0.741		
Total	212,485	100.000	6,332,177	100.000	1,271,370	2,871,878

Summary:
Calculated O2 (% wet):	3.01
Calculated O2 (% dry):	3.30
Calculated SO3 (ppm wel):	18
Calculated NOx (ppm wel):	230
Calculated NOx (ppm wet corr):	230
Calculated NOx (ppm dry):	252
Calculated NOx (ppm dry corr):	256
Calculated NOx (lb/MBTU):	0.35
Calculated Coal Feed (lb/hr):	532,000
Calculated Heat Input (MBTU/hr):	6,650

Exhibit I

700-MW New Plant - SCR Capital, O&M, and Levelized Costs for 60% Removal

Exhibit I

700 MW Plant - SCR Capital Cost for 60% Removal

Process Areas	k$	$/kw
Catalyst	$6,078	$8.7
Reactor Housing, Ductwork, Steel	$12,554	$17.9
Sootblowers	$1,202	$1.7
Ammonia Storage, Handling, and Injection	$1,549	$2.2
ID Fan Differential	$512	$0.7
Air Preheater Differential	$524	$0.7
Ash Handling Differential	$617	$0.9
Electrical	$319	$0.5
Instruments & Controls	$150	$0.2
Testing, Training, Commissioning	$175	$0.3
(A) Total Process Capital (sum of process areas)	$23,681	$33.8
(B) General Facilities (2% of A)	$474	$0.7
(C) Engineering (8% of A)	$1,894	$2.7
(D) Project Contingency (15% of A+B+C)	$3,907	$5.6
(E) Total Plant Cost (A+B+C+D)	$29,956	$42.8
(F) Allowance for Funds During Construction (1.91% of E)	$572	$0.8
(G) Total Plant Investment (E+F)	$30,528	$43.6
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$457	$0.7
(J) Inventory Capital	$342	$0.5
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$31,327	$44.8

Exhibit I

700 MW Plant - SCR Operating and Maintenance Cost for 60% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$189,000
Maintenance material				$284,000
Administration/support labor				$76,000
Subtotal Fixed Costs				$614,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	9.98	$2.00	$114,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	525	$0.13	$373,000
Catalyst	cu. ft.	(Note 1)	$400	$1,260,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	1788	$0.03	$306,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$2,053,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$2,667,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or replacement schedule. Dollar amount shown in this table represents a levelized annual reserve for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit I

700 MW Plant – Summary of Performance and Cost Data for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	658
Power produced, (net)	10^6 kWh/yr	3746.652
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	1,514,604
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	3848

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.254	0.116	0.970
Fixed O&M Cost	1.360	0.223	1.000	0.164
Variable O&M Cost	1.360	0.745	1.000	0.455
Total Cost		2.222		1.589
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,221	0.116	$944
Fixed O&M Cost	1.360	$218	1.000	$160
Variable O&M Cost	1.360	$726	1.000	$443
Total Cost		$2,165		$1,547

Exhibit J

250-MW New Plant
Summary of Performance and Levelized Cost Vs. Inlet NO$_X$ Concentration

Exhibit J

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 0.45 lb/MBTU Inlet NOx Concentration

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.45
Emission with controls	lb/MBTU	0.18
Amount removed	ton/yr	1767

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.788	1.000	0.485
Total Cost		2.611		1.882
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,139	0.116	$881
Fixed O&M Cost	1.362	$241	1.000	$177
Variable O&M Cost	1.362	$597	1.000	$367
Total Cost		$1,977		$1,425

Exhibit J

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 0.40 lb/MBTU Inlet NOx Concentration

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.40
Emission with controls	lb/MBTU	0.16
Amount removed	ton/yr	1571

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.767	1.000	0.469
Total Cost		2.590		1.866
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,281	0.116	$991
Fixed O&M Cost	1.362	$271	1.000	$199
Variable O&M Cost	1.362	$653	1.000	$400
Total Cost		$2,205		$1,590

Exhibit J

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 0.35 lb/MBTU Inlet NOx Concentration

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.746	1.000	0.454
Total Cost		2.569		1.851
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$726	1.000	$442
Total Cost		$2,500		$1,802

Exhibit J

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 0.30 lb/MBTU Inlet NOx Concentration

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced. (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.30
Emission with controls	lb/MBTU	0.12
Amount removed	ton/yr	1178

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.726	1.000	0.439
Total Cost		2.549		1.836
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,708	0.116	$1,321
Fixed O&M Cost	1.362	$362	1.000	$266
Variable O&M Cost	1.362	$824	1.000	$499
Total Cost		$2,894		$2,086

Exhibit J

250 MW Plant - 60% NOx Reduction
Summary-of-Performance and Cost Data
For 0.25 lb/MBTU Inlet NOx Concentration

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.25
Emission with controls	lb/MBTU	0.10
Amount removed	ton/yr	982

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	**Factor**	**Mills/kWh**	**Factor**	**Mills/kWh**
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.706	1.000	0.424
Total Cost		2.529		1.821
Levelized Cost, $/ton NOx Removed	**Factor**	**$/ton**	**Factor**	**$/ton**
Capital Charge	0.150	$2,050	0.116	$1,585
Fixed O&M Cost	1.362	$434	1.000	$319
Variable O&M Cost	1.362	$962	1.000	$579
Total Cost		$3,446		$2,483

Exhibit K

250-MW New Plant
Summary of Performance and Levelized Cost Vs. Catalyst Relative Activity
(Catalyst Management Plan)

Exhibit K

Catalyst Management Plan Sensitivity (K/Ko = 0.70)
250 MW Plant - SCR Operating and Maintenance Cost for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.891	1.000	0.535
Total Cost		2.714		1.932
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$868	1.000	$521
Total Cost		$2,642		$1,881

Exhibit K

Catalyst Management Plan Sensitivity (K/Ko = 0.80)
250 MW Plant - SCR Operating and Maintenance Cost for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.746	1.000	0.454
Total Cost		2.569		1.851
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$726	1.000	$442
Total Cost		$2,500		$1,802

Exhibit K

Catalyst Management Plan Sensitivity (K/Ko = 0.90)
250 MW Plant - SCR Operating and Maintenance Cost for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33
Emission Control Data	**Units**	**Value**
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	**Factor**	**Mills/kWh**	**Factor**	**Mills/kWh**
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.508	1.000	0.319
Total Cost		2.331		1.716
Levelized Cost, $/ton NOx Removed	**Factor**	**$/ton**	**Factor**	**$/ton**
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$495	1.000	$311
Total Cost		$2,269		$1,671

Exhibit L

250-MW New Plant
Summary of Performance and Levelized Cost Vs. Return on Common Equity

Exhibit L

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 7.0% Return on Common Equity

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.132	1.323	0.100	1.003
Fixed O&M Cost	1.395	0.326	1.000	0.234
Variable O&M Cost	1.395	0.708	1.000	0.454
Total Cost		2.357		1.691
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.132	$1,288	0.100	$976
Fixed O&M Cost	1.395	$318	1.000	$228
Variable O&M Cost	1.395	$689	1.000	$442
Total Cost		$2,295		$1,646

Exhibit L

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 9.0% Return on Common Equity

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.141	1.414	0.108	1.083
Fixed O&M Cost	1.378	0.322	1.000	0.234
Variable O&M Cost	1.378	0.727	1.000	0.454
Total Cost		2.463		1.771
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.141	$1,376	0.108	$1,054
Fixed O&M Cost	1.378	$314	1.000	$228
Variable O&M Cost	1.378	$708	1.000	$442
Total Cost		$2,398		$1,724

Exhibit L

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 11.0% Return on Common Equity

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.504	0.116	1.163
Fixed O&M Cost	1.362	0.319	1.000	0.234
Variable O&M Cost	1.362	0.746	1.000	0.454
Total Cost		2.569		1.851
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,464	0.116	$1,132
Fixed O&M Cost	1.362	$310	1.000	$228
Variable O&M Cost	1.362	$726	1.000	$442
Total Cost		$2,500		$1,802

Exhibit L

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 13.0% Return on Common Equity

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.160	1.604	0.124	1.243
Fixed O&M Cost	1.347	0.315	1.000	0.234
Variable O&M Cost	1.347	0.766	1.000	0.454
Total Cost		2.685		1.931
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.160	$1,562	0.124	$1,210
Fixed O&M Cost	1.347	$307	1.000	$228
Variable O&M Cost	1.347	$746	1.000	$442
Total Cost		$2,615		$1,880

Exhibit L

250 MW Plant - 60% NOx Reduction
Summary of Performance and Cost Data
For 15.0% Return on Common Equity

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.169	1.694	0.133	1.333
Fixed O&M Cost	1.333	0.312	1.000	0.234
Variable O&M Cost	1.333	0.786	1.000	0.454
Total Cost		2.792		2.021
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.169	$1,650	0.133	$1,298
Fixed O&M Cost	1.333	$304	1.000	$228
Variable O&M Cost	1.333	$766	1.000	$442
Total Cost		$2,720		$1,968

Exhibit M

250-MW New Plant
Summary of Capital, O&M, and Levelized Cost Vs. Catalyst Price

Exhibit M

Catalyst Price Sensitivity ($350/ft3)
250 MW Base Case SCR Capital Cost for 60% Removal

Process Areas	k$	$/kw
Catalyst (@ $350/ft3)	$1,897	$7.6
Reactor Housing, Ductwork, Steel	$4,958	$19.8
Sootblowers	$580	$2.3
Ammonia Storage, Handling, and Injection	$1,292	$5.2
ID Fan Differential	$216	$0.9
Air Preheater Differential	$220	$0.9
Ash Handling Differential	$300	$1.2
Electrical	$201	$0.8
Instruments & Controls	$100	$0.4
Testing, Training, Commissioning	$138	$0.6
(A) Total Process Capital (sum of process areas)	$9,901	$39.6
(B) General Facilities (2% of A)	$198	$0.8
(C) Engineering (8% of A)	$792	$3.2
(D) Project Contingency (15% of A+B+C)	$1,634	$6.5
(E) Total Plant Cost (A+B+C+D)	$12,525	$50.1
(F) Allowance for Funds During Construction (1.91% of E)	$239	$1.0
(G) Total Plant Investment (E+F)	$12,764	$51.1
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$165	$0.7
(J) Inventory Capital	$111	$0.4
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$13,040	$52.2

Exhibit M

Catalyst Price Sensitivity   (@ $350/ft3)
250 MW Base Case SCR Operating and Maintenance Cost for 60% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$79,000
Maintenance material				$119,000
Administration/support labor				$43,000
Subtotal Fixed Costs				$306,000
Variable Operating Costs	**Units**	**Quantity**	**$/Unit**	**$/ yr**
Fuels				
Coal	MBTU/hr	3.56	$2.00	$41,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	187	$0.13	$133,000
Catalyst	cu. ft.	(Note 1)	$350	$394,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	639	$0.03	$109,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$677,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$983,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or
replacement schedule. Dollar amount shown in this table represents a levelized annual reserve
for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit M

Catalyst Price Sensitivity (@ $350/ft3)
250 MW Base Case SCR Operating and Maintenance Cost for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.462	0.116	1.130
Fixed O&M Cost	1.362	0.312	1.000	0.229
Variable O&M Cost	1.362	0.689	1.000	0.424
Total Cost		2.463		1.783
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,423	0.116	$1,101
Fixed O&M Cost	1.362	$304	1.000	$223
Variable O&M Cost	1.362	$671	1.000	$413
Total Cost		$2,398		$1,737

Exhibit M

Catalyst Price Sensitivity ($450/ft3)
250 MW Base Case SCR Capital Cost for 60% Removal

Process Areas	k$	$/kw
Catalyst (@ $450/ft3)	$2,438	$9.8
Reactor Housing, Ductwork, Steel	$4,958	$19.8
Sootblowers	$580	$2.3
Ammonia Storage, Handling, and Injection	$1,292	$5.2
ID Fan Differential	$216	$0.9
Air Preheater Differential	$220	$0.9
Ash Handling Differential	$300	$1.2
Electrical	$201	$0.8
Instruments & Controls	$100	$0.4
Testing, Training, Commissioning	$138	$0.6
(A) Total Process Capital (sum of process areas)	$10,443	$41.8
(B) General Facilities (2% of A)	$209	$0.8
(C) Engineering (8% of A)	$835	$3.3
(D) Project Contingency (15% of A+B+C)	$1,723	$6.9
(E) Total Plant Cost (A+B+C+D)	$13,210	$52.8
(F) Allowance for Funds During Construction (1.91% of E)	$252	$1.0
(G) Total Plant Investment (E+F)	$13,463	$53.9
(H) Royalty Allowance (0% of A)	$0	$0.0
(I) Preproduction Cost (2 month startup)	$184	$0.7
(J) Inventory Capital	$130	$0.5
K) Initial Catalyst and Chemical	$0	$0.0
(L) Total Capital Requirements (G+H+I+J+K)	$13,777	$55.1

Exhibit M

Catalyst Price Sensitivity   (@ $450/ft3)
250 MW Base Case SCR Operating and Maintenance Cost for 60% Removal

Fixed O&M Costs	Units	Quantity	$/Unit	$/ yr
Operating labor	Man-hr	2847	$23.00	$65,000
Maintenance labor				$84,000
Maintenance material				$125,000
Administration/support labor				$45,000
Subtotal Fixed Costs				$319,000

Variable Operating Costs	Units	Quantity	$/Unit	$/ yr
Fuels				
Coal	MBTU/hr	3.56	$2.00	$41,000
Sorbent				
n/a				$0
Chemicals/Catalyst				
Ammonia	lb/hr	187	$0.13	$133,000
Catalyst	cu. ft.	(Note 1)	$450	$506,000
Utilities				
Condensate	10^3 lb/hr			$0
Raw water	10^3 gal/hr			$0
Cooling water	10^3 gal/hr			$0
LP steam (0-70 psia)	10^3 lb/hr			$0
MP steam (70-250 psia)	10^3 lb/hr			$0
HP steam (>250 psia)	10^3 lb/hr			$0
Electric power	kWh/hr	639	$0.03	$109,000
Byproduct Credits				
n/a				$0
Waste Disposal Charges				
n/a				$0
Subtotal Variable Cost				$789,000
TOTAL O&M COSTS (FIXED + VARIABLE)				$1,108,000

Note 1 - Catalyst is not replaced on a yearly basis. Refer to catalyst management plan for addition and/or replacement schedule. Dollar amount shown in this table represents a levelized annual reserve for replacement based on present worth analysis of the catalyst replacement schedule.

Exhibit M

Catalyst Price Sensitivity   (@ $450/ft3)
250 MW Base Case SCR Operating and Maintenance Cost for 60% Removal

Power Plant Attributes	Units	Value
Plant capacity, (net)	MWe	235
Power produced, (net)	10^6 kWh/yr	1338.09
Capacity factor	%	65
Plant life	years	30
Coal feed	tons/yr	540,930
Sulfur in coal	wt %	2.33

Emission Control Data	Units	Value
SCR removal efficiency	%	60
Emission without controls	lb/MBTU	0.35
Emission with controls	lb/MBTU	0.14
Amount removed	ton/yr	1374

	Current Dollars		Constant Dollars	
Levelized Cost, mills/kWhr	Factor	Mills/kWh	Factor	Mills/kWh
Capital Charge	0.150	1.544	0.116	1.194
Fixed O&M Cost	1.362	0.325	1.000	0.238
Variable O&M Cost	1.362	0.803	1.000	0.485
Total Cost		2.672		1.917
Levelized Cost, $/ton NOx Removed	Factor	$/ton	Factor	$/ton
Capital Charge	0.150	$1,504	0.116	$1,163
Fixed O&M Cost	1.362	$316	1.000	$232
Variable O&M Cost	1.362	$782	1.000	$472
Total Cost		$2,602		$1,867